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Two-Stage Biological Treatment of Coke Plant Wastewater

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TWO-STAGE BIOLOGICAL TREATMENT OF COKE PLANT WASTEWATER

Executive Summary

by

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16. ABSTRACT The report documents a pilot-plant study of the use of advanced waste treatment methods in upgrading metallurgical cokemaking wastewaters to Best Available Technology (BAT) levels. Mobile treatment units, operable at a flow rate of 19 l/min, were used. Methods used included two-stage activated sludge treatment for removal of organic carbon compounds and ammonia; filtration and activated carbon were also studied as polishing steps. For each treatment studied, samples (including toxic pollutants) and operational data were obtained for later use in assessing and comparing treatment adequacy. The study showed that high levels of organic pollutant removal were achieved in the 1st stage of biological oxidation. Abnormal operating conditions in the coke plant wastewater pretreatment system during the testing resulted in ammonia levels as high as 2000 mg/l, making it necessary to dilute the 2nd stage feed before nitrification could be achieved. Ammonia reductions of >97% were achieved in the 2nd stage. Activated carbon and filtration effectively removed suspended solids, total organic carbon, color, and thiocyanate.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Carbon Compounds	Pollution Control	13B	07C
Waste Water	Organic Compounds	Stationary Sources		
Coking	Ammonia	Biological Oxidation	13H	07B
Biochemical Oxygen Demand	Filtration			07D
	Activated Carbon		06C	11G
Sludge	Decoloring		07A	

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SECTION 1

INTRODUCTION

PROJECT PURPOSE

The primary purpose of this project was to investigate the technical and economical feasibility of biological treatment of by-product cokemaking wastewater to Best Available Technology (BAT) levels.* Principal wastewater sources produced from the by-product recovery process at the plant investigated included excess ammonia liquor and benzol plant wastes. Wastewater treatment at this plant consisted of stripping the waste ammonia liquor via steam and caustic solutions, followed by blending with the benzol plant wastes and biological treatment with a single stage activated sludge system. The wastewater for the pilot study was obtained after an equalization basin and prior to the existing biological system. The combined wastewater contained pollutants such as suspended solids, oil and grease, ammonia, phenolic compounds, cyanide, sulfide, thiocyanates, and toxic materials.

The study was performed using the mobile bio-oxidation system of the U.S. Environmental Protection Agency (EPA). This pilot plant is contained in a semi-trailer van as shown in Figure 1. Near the end of the investigation, a second EPA semi-trailer van containing the activated carbon pilot plant system (Figure 2) was brought to the test site. The purpose of this system was to evaluate the efficiency of activated carbon for removing toxic pollutants present in the coke plant wastewater. A small pilot-size filtration unit was investigated as a method of polishing the biological oxidation effluent.

The wastewater treatment trains that were investigated on a pilot scale included the following:

1. AS(C) + AS(N)
2. AS(C) + AS(N) + AC

AS(C): Activated sludge (carbonaceous removal)

---

*Note: Because the development of final effluent Limitation Guidelines was underway during the time of this study, the reference BPT and BAT pollutant parameters used throughout this report are those listed in the "Development Document for Proposed Effluent Limitations Guidelines and Standards for the Iron and Steel Manufacturing Point Source Category - Vol. II, By-Product Cokemaking Subcategory, October, 1979, ERA-4401 1-79/02400 (1)".

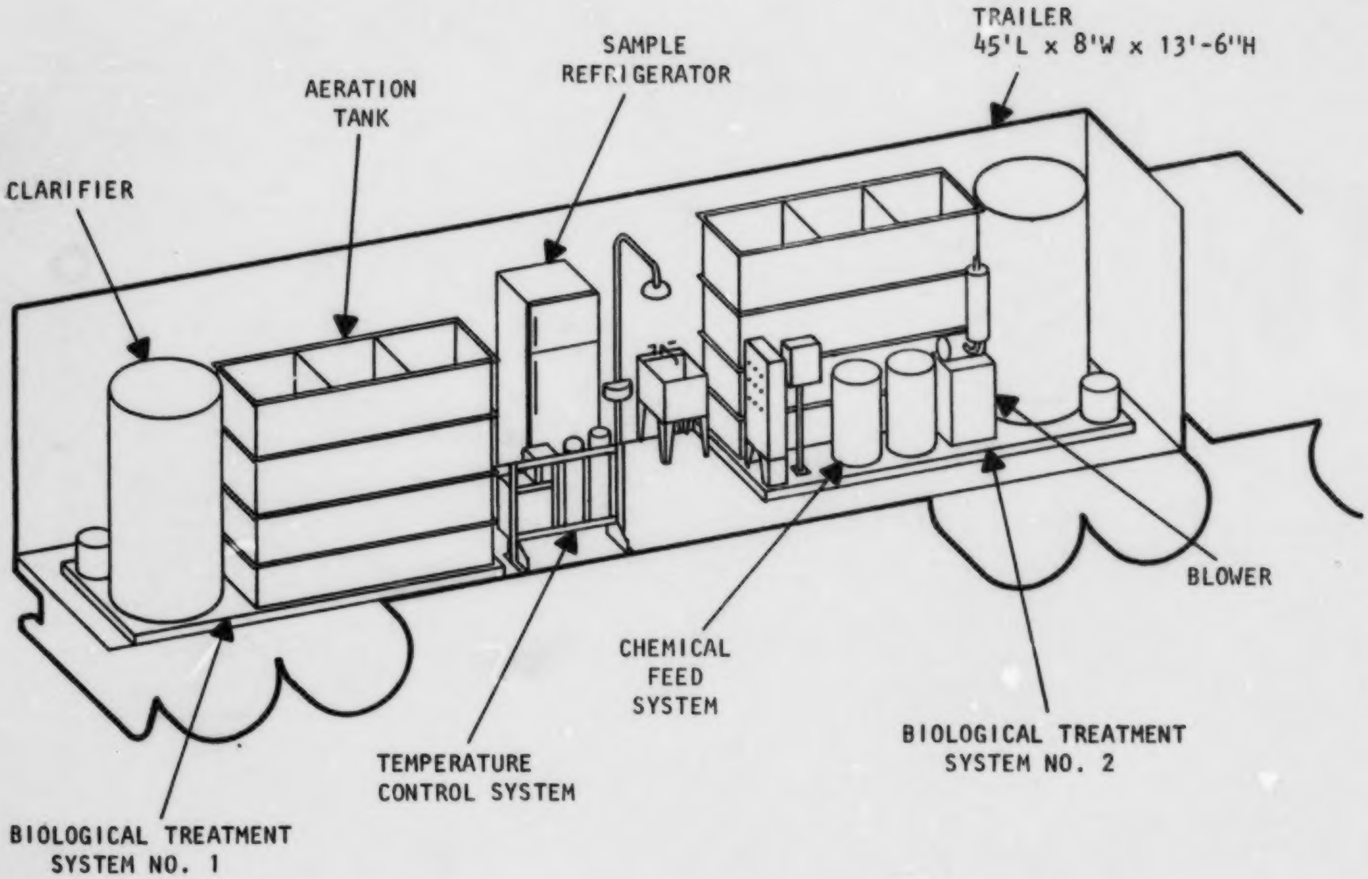

TRAILER
45'L x 8'W x 13'-6"H
SAMPLE
REFRIGERATOR
AERATION
TANK
CLARIFIER
BLOWER
CHEMICAL
FEED
SYSTEM
BIOLOGICAL TREATMENT
SYSTEM NO. 2
TEMPERATURE
CONTROL SYSTEM
BIOLOGICAL TREATMENT
SYSTEM NO. 1

Figure 1. Steel plant mobile bio-oxidation treatment system-trailer no. 3.

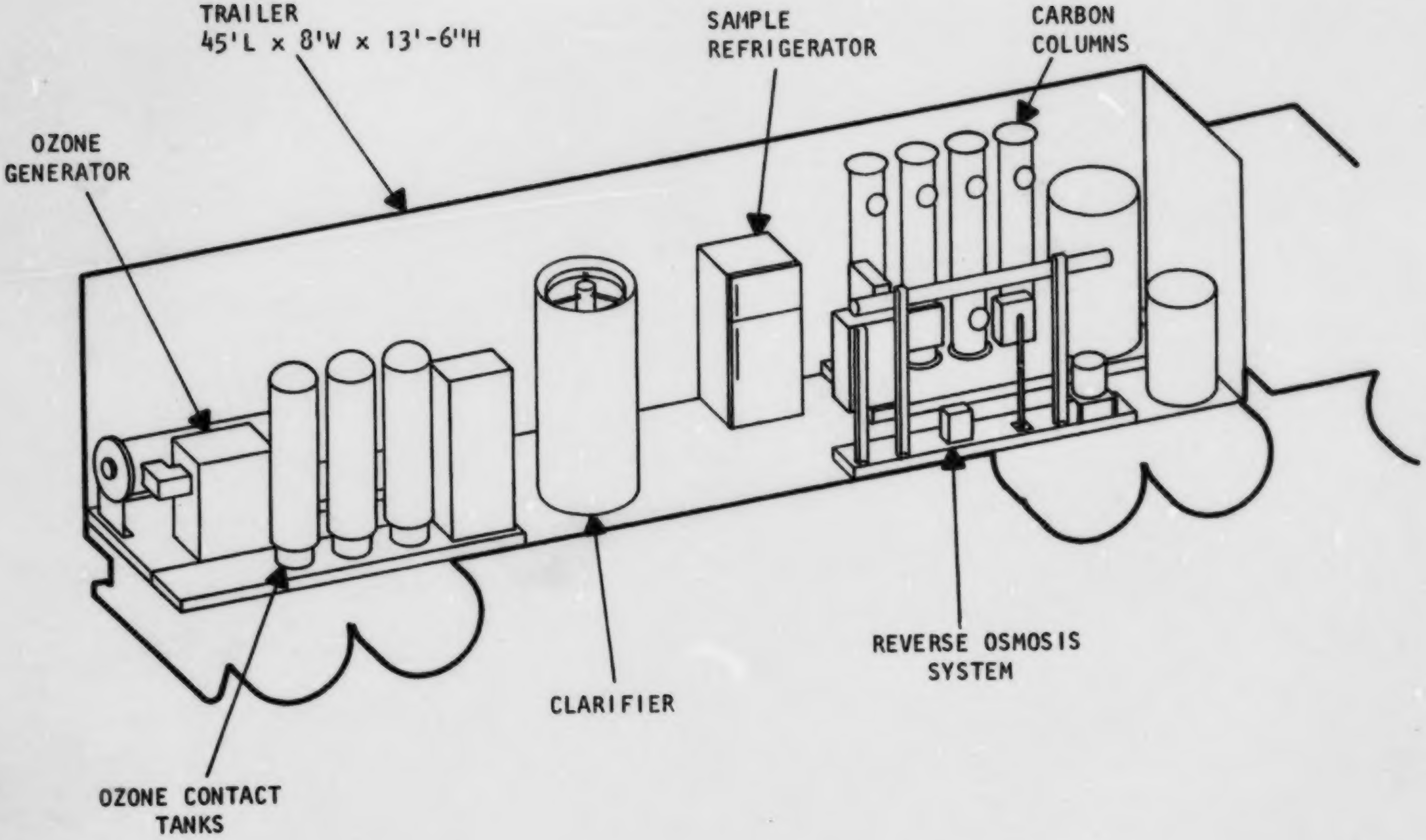

TRAILER
45'L x 8'W x 13'-6"H
SAMPLE
REFRIGERATOR
CARBON
COLUMNS
OZONE
GENERATOR
OZONE CONTACT
TANKS
CLARIFIER
REVERSE OSMOSIS
SYSTEM

Figure 2. Steel plant mobile treatment system-trailer no. 2.

3. AS(C) + AS(N) + DMF

AS(N): activated sludge (nitrification)
AC: activated carbon
DMF: dual media filtration

In the first treatment train, plant wastewater from downstream of the coke plant cooling tower was passed through a mixing tank, through the first stage activated sludge system (carbonaceous removal) and then through the second stage activated sludge system (nitrogen removal). The second treatment train consisted of the first treatment train (bio-oxidation) followed by activated carbon adsorption. The third treatment train included the components of the first treatment train followed by dual media filtration. The detailed treatment train arrangements are shown graphically in Figure 3.

The small sump tank shown in Figure 3 was used for pH adjustment, dilution, and chemical dispersal. As the wastewater flowed through the first stage activated sludge system, carbonaceous material (BOD, phenol, etc.) was removed. Effluent from the first stage clarifier was pumped through the second stage activated sludge system where ammonia nitrogen was oxidized by the nitrification process. The use of powdered activated carbon was investigated for removal of toxics and to improve settling. Final effluent from the second stage activated sludge system was passed through the activated carbon or dual media filtration systems late in the test program to complete the second and third treatment train arrangements respectively.

During the pilot study, samples were collected for both conventional and toxic pollutants. Operational data were also recorded on a daily basis to evaluate the effectiveness of the respective treatment trains. The pilot study results were evaluated using the three primary criteria listed below:

1. process and/or treatment train performance
2. capital and operating costs
3. space requirements

The forthcoming sections of this report discuss in detail the findings of this pilot scale investigation. An individual section has been prepared for each technology investigated to discuss results, operational problems, and full-scale system costs and space requirements. Section 2 combines the individual process findings into schemes capable of treating the coke plant wastewater to acceptable levels. Raw data obtained during the pilot study are included as part of the Appendix.

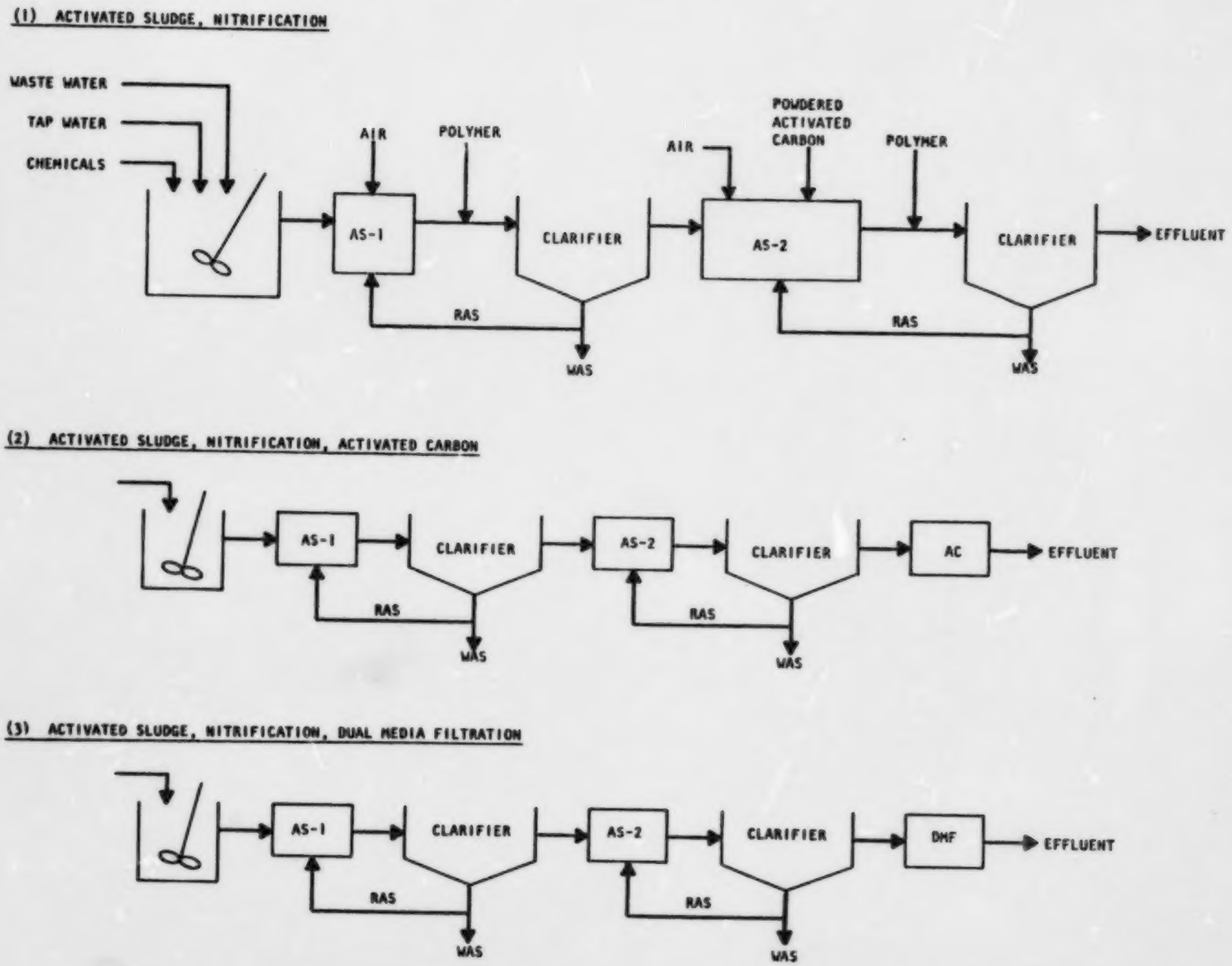

(1) ACTIVATED SLUDGE, NITRIFICATION
WASTE WATER
TAP WATER
CHEMICALS
AIR
POLYMER
AS-1
CLARIFIER
RAS
WAS
AIR
POWDERED ACTIVATED CARBON
POLYMER
AS-2
CLARIFIER
RAS
WAS
EFFLUENT
(2) ACTIVATED SLUDGE, NITRIFICATION, ACTIVATED CARBON
AS-1
CLARIFIER
RAS
WAS
AS-2
CLARIFIER
RAS
WAS
AC
EFFLUENT
(3) ACTIVATED SLUDGE, NITRIFICATION, DUAL MEDIA FILTRATION
AS-1
CLARIFIER
RAS
WAS
AS-2
CLARIFIER
RAS
WAS
DMF
EFFLUENT

Figure 3. Process trains investigated for treatment of by-product coke plant wastewater.

SECTION 2

SUMMARY AND EVALUATION

EXPERIMENTAL RESULTS

General

The biological investigation on the coke plant's by-product coke-making wastewater was conducted between October 16, 1979 and February 14, 1980. During the study, coke production averaged 4,372 metric tons (4,816 tons) per day, while the average wastewater flow was 1,885 m^3/day (0.499 mgd). The corresponding water application rate (liters of water/metric tons of coke produced) during the study was 432 liters/kkg (104 gal./ton). Based on BAT (non-priority pollutant) limits (1), the allowable pollutant concentrations in the effluent would be:

Parameter	BAT limit
pH	6.0-9.0
Ammonia, mg/l	32
Cyanide-T, mg/l	2
Oil and Grease, mg/l	20

Parameter	BAT limit
Phenol, mg/l	0.11
Sulfide, mg/l	0.6
Suspended Solids, mg/l	86
Thiocyanate, mg/l	1

The pilot studies were structured toward meeting the above guidelines.

Three pilot scale treatment trains were investigated:

1. Two stage activated sludge.
2. Two stage activated sludge followed by activated carbon.
3. Two stage activated sludge followed by dual media filtration.

The capital cost information presented in this report has been prepared from a minimum of flow sheet and equipment data. Using the definition found in Perry's "Chemical Engineers' Handbook" (2), this type of estimate would be termed a "study estimate", having an accuracy of ±30 percent.

Wastewater Treatment System

There are two primary process water streams associated with the by-product cokemaking operations at the plant investigated: namely, 1) benzol plant effluent and 2) ammonia still excess liquor.

The benzol plant effluent is blended with the ammonia still excess liquor and the combined waste is equalized and passed through a cooling tower. The waste is diluted with coal yard runoff and service water and undergoes biological treatment. The dilution process consists of blending 3 parts of coke plant wastewater with 1 part of service water. After biological treatment, the waste is clarified and discharged to the Ohio River. Feedwater to the mobile system was taken from a pressurized line prior to dilution for the entire duration of the study. This undiluted waste was fed to the pilot system's small sump where dilution water was added in the ratio similar to the full-scale system (1 part dilution water to 3 parts coke plant wastewater).

Characteristics of the wastewater obtained from the pilot system's small sump during the study are shown in Table 1.

TABLE 1. COMPARISON OF ACTUAL PLANT VALUES OBSERVED DURING THE STUDY PERIOD TO EXPECTED VALUES WITH EQUIVALENT PRETREATMENT SYSTEM.[a]

Parameter	Expected Values[a]	Analytical value from the research site[b] Average	Range
Cyanide T, mg/l	30	9.9	0.8-32
Phenol, mg/l	2	655	440-920
Ammonia, mg/l	125	772	220-2,090
Oil and Grease, mg/l	15	17	2-105
Suspended Solids, mg/l	50	69	20-205
pH	6.0-9.0	--	6.80-8.70

a. Fixed and free ammonia still treatment.

b. Diluted 3 parts coke plant wastewater with 1 part service water.

As shown in the table, concentrations of phenol and ammonia in the pilot system feedwater were well above the expected values during the study. The pollutant values shown remained high and quite variable throughout the test period. The plant personnel indicated that this fact was due to abnormal operation of the ammonia stills. This abnormal waste stream greatly diminished the usefulness of the study results, creating artificial operating conditions that would not be encountered in actual full scale practice. Nonetheless, an effort has been made to gain as much knowledge from the results as possible and to apply the results to expected actual practice where possible.

First Stage Activated Sludge (Carbonaceous Removal) Results

Performance Data--

The first stage pilot activated sludge system was effective in removing both BOD and phenol. Influent BOD ranged from 1,290 mg/l to 2,550 mg/l and averaged 1,800 mg/l. The BOD removal efficiency was typically 95 percent with a range from 60 percent to 99 percent. Good removals of phenol were also observed in the first stage system. Efficiency ranged from 90 percent to 100 percent. The average diluted feed phenol concentration was 655 mg/l, with a range from 440 mg/l to 920 mg/l. Removals of thiocyanate and sulfide were also observed as seen in Table 2.

Operating Data--

Operating data used in controlling the first stage activated sludge unit included sludge volume index (SVI), mixed liquor volatile suspended solids, settling rate and food to microorganism (F/M) ratio.

With regard to settling, it was necessary to add a high strength cationic polymer to improve mixed liquor settleability in the first stage clarifier. A dosage of about 5 mg/l would be recommended for full-scale application. The settling rate varied from 0.003 m/min (0.01 ft/min) to 0.07 m/min (0.22 ft/min) and averaged 0.03 m/min (0.09 ft/min). The sludge volume index during the study averaged 87 and ranged from 18-155. A SVI of less than 100 is indicative of a good settling sludge.

During the initial start-up period, food to microorganism ratio was quite variable as shown in Figure 4. However, after stabilization, the F/M remained consistent and good pollutant removals were achieved.

TABLE 2. INFLUENT VERSUS EFFLUENT VALUES - UNIT AS-1

Parameter	AS-1 Influent[a]	AS-1 Effluent
pH	7.55[b]	7.10[b]
Suspended Solids, mg/l	69	127
Sulfide, mg/l	31	5
Phenol, mg/l	655	8.3
Oil and Grease, mg/l	17	10
Ammonia, mg/l	772	720
Cyanide-T, mg/l	9.9	9.2
BOD, mg/l	1,800	215
Thiocyanate, mg/l	450	323

a. Influent consisted of 3 parts coke plant wastewater + 1 part service water.

b. Typical value.

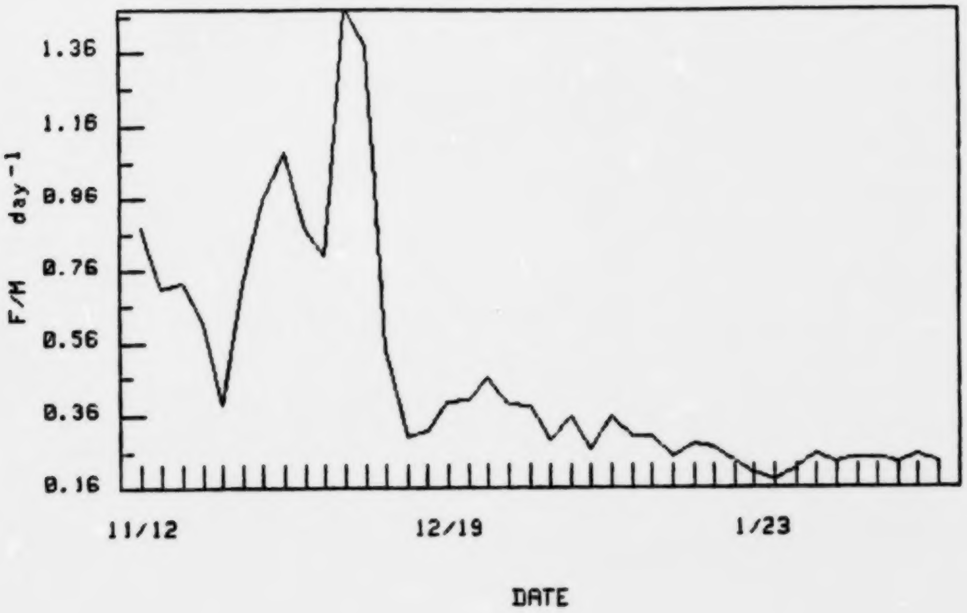


Figure 4. F/M ratio - unit AS-1.

Another important operating consideration was mixed liquor volatile suspended solids (MLVSS), since selection of this concentration directly affects other process parameters such as F/M ratio, hydraulic retention time, sludge production and sludge age. During the pilot study, operating at a MLVSS concentration of about 3,500 mg/l provided optimum treatment.

Ultimately, four periods were selected during which consistent operation of the pilot system was maintained. A summary of these periods is shown is Table 3.

TABLE 3. SELECTED PERIODS OF AS-I CONSISTENT OPERATION

Period	Ave. F/M, day^{-1}	Ave. BOD removal, %	Ave. SVI	Ave. settling rate, ft/min
11/23-11/28	0.917	83.4	64	0.067
11/12-11/19	0.735	71.3	120	0.062
12/19-12/26	0.413	90.0	48	0.036
1/30-2/7	0.24	93.0	106	0.033

Note: ft/min x 0.3048 = m/min

The information provided in this table was used for scale-up to a full-size installation.

Gravity thickening tests were also performed on the settled sludge. It was concluded that a full-scale gravity thickening system could achieve an underflow suspended solids concentration of 20,000 mg/l while operating at a mass loading of 11 $kg/m^2/day$ (2.25 $lb/ft^2/day$).

Cost Data

Listed in Tables 4 and 5 are the costs associated with treatment of the coke plant wastewater by first stage activated sludge. In development of the cost data, an equalized feed flow has been assumed. Cost estimates have been prepared for design flows of 379 m^3/day (0.1 mgd), 1,892 m^3/day (0.5 mgd) and 3,785 m^3/day (1.0 mgd). Cost data were developed for treatment of the coke plant wastewater as received (Design 1 - Table 4) and for an assumed coke plant feedwater (Design 2 - Table 5) that would result from proper still operation. Dilution water would be required in Design 1 and not in Design 2.

Comparing the cost data in Tables 4 and 5, it is evident that cost savings of approximately 10 percent in capital investment and 7 percent in operational expenses would result if the wastewater were pretreated properly in the ammonia stills. For example, assuming current wastewater pollutant concentrations, an estimated treatment cost of \$4.51/3,785 liters (\$4.51/1,000 gal.) would be realized for a 1,893 m^3/day (0.5 mgd) wastewater treatment facility. By reducing effluent concentrations by proper ammonia still operation, this cost could be reduced to \$4.21/3,785 liters (\$4.21/1,000 gal.). Both systems were designed for an organic removal of 90 percent.

TABLE 4. FIRST STAGE ACTIVATED SLUDGE
COST SUMMARY - DESIGN 1[a]

Size, m^3/day (mgd)	379 (0.1)	1,893 (0.5)	3,785 (1.0)
Capital costs, (\$)[b]	838,000	2,654,000	4,373,000
Annualized costs, (\$/yr)[b,c]	233,000	823,000	1,495,000
Annualized costs, \$/3,785 ℓ (\$/1,000 gal.)[b,c]	6.38	4.51	4.10
Land requirement, m^2 (ft^2)	440 (4,700)	1,200 (12,900)	3,600 (38,400)

a. Includes cost of chemical addition system; dilution ratio of 3 parts coke plant wastewater to 1 part service water assumed.

b. All prices are effective as of December, 1980.

c. Annualized cost = operating cost + capital amortization (20 yr. life; 7.125%).

TABLE 5. FIRST STAGE ACTIVATED SLUDGE
COST SUMMARY - DESIGN 2[a]

Size, m^3/day (mgd)	379 (0[illegible]	1,893 (0.5)	3,785 (1.0)
Capital costs, (\$)[b]	792,000	2,427,000	3,965,000
Annualized costs,(\$/yr)[b,c]	156,000	559,000	1,040,000
Annualized costs, \$/3,785 ℓ (\$/1,000 gal.)[b,c]	6.05	4.21	3.81
Land requirement, m^2 (ft^2)	240 (2,600)	1,700 (18,700)	3,200 (34,600)

a. Includes cost of pretreatment system.

b. All prices are effective as of December, 1980.

c. Annualized cost = operating cost + capital amoritization (20 yr. life; 7.125%).

Similarly, land requirements would be significantly reduced for the second design. It is estimated that about a 45 percent savings of space would result if the coke plant wastewater were pretreated by proper ammonia still operation.

Second Stage Activated Sludge (Nitrogen Removal) Results

Performance Data--

The primary objective of the second stage bioxidation unit was to reduce influent concentrations of ammonia through the nitrification process. The pilot system feed ammonia during the study was quite variable, ranging from 230 mg/l to 2,090 mg/l after dilution (this refers to dilution of the first stage feedwater - 3 parts coke plant wastewater to 1 part service water). The average diluted feed ammonia concentration to the pilot units was 772 mg/l. The influent wastewater to the second stage was pH adjusted using sodium carbonate and sodium hydroxide. Powdered activated carbon was added to reduce the effect of toxic shock loads and help weight the nitrified sludge. Polymer was also fed to the second stage clarifier to minimize biomass losses over the weir.

Attempts to achieve a substantial population of nitrifiers were unsuccessful during the first six weeks of testing. Dilution to the second stage was initiated on December 1, 1979 to reduce the feed ammonia concentration and therefore promote the growth of nitrifying bacteria. (The first stage effluent was diluted to reduce second stage influent ammonia concentrations from 772 mg/l down to about 200 mg/l.) This had a positive effect on the microorganism population and nitrification began to take place in early January, 1980. Excellent ammonia removals were then consistently achieved for the duration of the project as shown in Figure 5.

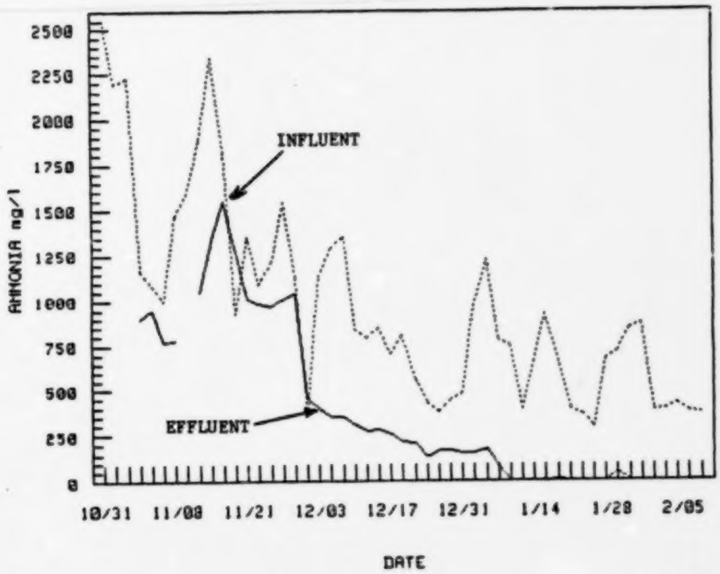


Figure 5. Influent ammonia versus AS-II effluent ammonia.

Test results for the period of time during which nitrification was occurring are presented in Table 6. During this period, ammonia removal averaged 90.6 percent. There was also some removal of thiocyanate and sulfide.

TABLE 6. INFLUENT AND EFFLUENT QUALITY FOR UNIT AS-II DURING ACTIVE NITRIFICATION (JANUARY 9 - FEBRUARY 7, 1980)

Parameter	AS-II Influent Ave.	AS-II Influent Range	AS-II Effluent Ave.	AS-II Effluent Range
SS, mg/l	45	10-95	155	25-340
S, mg/l	0.9	0.1-1.8	I	-
Phenol, mg/l	0.3	<0.1-0.7	0.3	<0.1-0.9
O&G, mg/l	3	<1-7	9	3-21
TKN, mg/l	183	78-267	25	8-73
NH_3 as N, mg/l	160	68-234	15	1-45
CN_T, mg/l	5.0	1.5-13.3	8.2[a]	2.1-30
CN_A, mg/l	0.49	<0.05-3.1	0.57	<0.05-5.1
TOC, mg/l	68	28-190	73	21-158
BOD, mg/l	30	4-97	25	4-87
SCN, mg/l	65	<1-180	5	<1-65
NO_2/NO_3 as N, mg/l	<1	-	186	160-215

I: Denotes Interference
a: Nitrite interference suspected. See Section 6 for complete discussion.

Notice from Table 6 that a decrease in effluent ammonia resulted in an increase in effluent nitrite-nitrate. During the nitrification period, effluent nitrite-nitrate averaged 186 mg/l. As nitrification took place, it was also evident that total cyanide effluent values began to increase and actually exceeded the influent concentrations. A literature review was performed and it was determined that the high effluent cyanide values were caused by interference. A complete discussion of this is presented in Section 6.

During nitrification, TKN was reduced from 183 mg/l to 29 mg/l, a reduction of 86 percent. There was no noticeable reduction in oil and grease or phenol.

Because of the high dilution of the waste stream, it was difficult to draw conclusive results as to the ability of AS-II to meet BAT requirements. However, significant reductions in pollutant concentrations were achieved. Ammonia concentrations were reduced from an average of 160 mg/l to 15 mg/l; the inlet ammonia level of 160 mg/l would be expected from a well-operated ammonia still system prior to AS-II. However, because this influent ammonia level was artifically met using dilution water in an amount one to three times the amount of wastewater, conclusions related to the effect of other pollutant parameters on the ability of AS-II to achieve a high degree of nitrification could not be determined.

Operating Data --
Sludge volume index during the study was typically less than 60, which is indicative of a good settling sludge. To improve settling rate, a high charge cationic polymer was added to the mixed liquor prior to clarification. An approximate dosage of 5 mg/l was determined from the pilot study. Settling rate ranged from 0.002 m/min (0.008 ft/min) to 0.14 m/min (0.47 ft/min).

The average mixed liquor settling rate was 0.043 m/min (0.14 ft/min). During the period of nitrification, mixed liquor suspended solids ranged from 6,000 mg/l to 9,000 mg/l. The F/M ratio was stable during nitrification, averaging 0.023 day^{-1} with a range between 0.017 day^{-1} and 0.028 day^{-1}.

For purposes of design, three operating periods were selected during the study which had uniform F/M ratios and these are shown in Table 7.

TABLE 7. SELECTED INTERVALS FOR UNIT AS-II

Period	Ave. F/M, day^{-1}	Ave. NH_3 removal, %	Ave. SVI	Ave. settling rate, ft/min
1/9-1/14/80	0.014	78.8	55	0.10
1/16-23/80	0.024	96.6	43	0.16
1/30-2/7/80	0.025	89.2	40	0.22

Note: ft/min x 0.3048 = m/min

The data provided in this table were used to scale up the pilot study findings to full-scale application. Late in the investigation, gravity thickening tests were conducted. Bench scale testing determined that a full-scale gravity thickener could obtain an underflow suspended solids concentration of 32,000 mg/l. The corresponding mass loading rate would be 21.6 kg/m^2/day (4.42 lb/ft^2/day).

Cost Data--

A summary of the costs associated with the second stage activated sludge treatment of the coke plant wastewater is shown in Tables 8 and 9.

TABLE 8. SECOND STAGE ACTIVATED SLUDGE COST SUMMARY - WASTEWATER AS RECEIVED[a]

Size, m^3/day (mgd)	379 (0.1)	1,893 (0.5)	3,785 (1.0)
Capital costs, (\$)[b]	989,000	3,209,000	5,784,000
Annualized costs, (\$/yr)[b,c]	590,000	2,587,000	5,063,000
Annualized costs,[b,c] \$/3,785 ℓ (\$/1,000 gal.)	16.16	14.17	13.87
Land requirement, m^2 (ft^2)	1,210 (13,000)	5,400 (58,200)	9,100 (97,500)

a. Assuming influent ammonia of 700 mg/l is diluted to 200 mg/l. Plant size shown is before dilution.

b. All prices are effective as of December, 1980.

c. Annualized cost = operating cost + capital amortization (20 yr. life; 7.125%).

TABLE 9. SECOND STAGE ACTIVATED SLUDGE
COST SUMMARY - PROPERLY PRETREATED FEEDWATER ASSUMED[a]

Size, m^3/day (mgd)	379 (0.1)	1,893 (0.5)	3,785 (1.0)
Capital costs, ($) [b]	518,000	1,238,000	2,222,000
Annualized costs, ($/yr)[b,c]	211,000	751,000	1,425,000
Annualized costs, $/3,785 ℓ ($/1,000 gal.)[b,c]	5.78	4.12	3.90
Land requirement, m^2 (ft^2)	160 (1,800)	1,550 (16,700)	2,750 (29,600)

a. Treated to an ammonia level of 200 mg/l.
b. All prices are effective as of December, 1980.
c. Annualized cost = operating cost + capital amortization (20 yr. life; 7.125%).

As shown in Tables 8 and 9, both designs were prepared for coke plant wastewater flows of 379 m^3/day (0.1 mgd), 1,893 m^3/day (0.50 mgd) and 3,785 m^3/day (1.0 mgd). The cost data presented in Table 8 assume the addition of dilution water. A dilution ratio of 3.5 parts service water to 1.0 part of coke plant wastewater was assumed. The Table 9 cost data assume good pretreatment with free and fixed ammonia stills so that no dilution water is required. Comparison of Tables 8 and 9 indicates that a capital savings of about 259 percent would result if the coke plant could provide better pretreatment of the wastewater to ammonia levels of 200 mg/l. Also annualized costs could be reduced by over 300 percent. For example, the annualized cost to treat the current wastewater discharge [1,893 m^3/day (0.5 mgd)] is estimated at $14.17/3,785 liters ($14.17/1,000 gal.). If the wastewater were better pretreated to an ammonia level of 200 mg/l, the expected annualized cost would drop to $4.12/3,785 liters ($4.12/1,000 gal.).

Similarly, land requirements would be less for a system designed to meet BAT guidelines if good fixed and free ammonia still operation were practiced.

Activated Carbon Results

The results of the biological test site activated carbon study determined that carbon was effective in removing TOC (53% removal), color (60% removal), BOD (40%) and SCN (79% removal) from the second stage activated sludge effluent. There were also minor removals of phenol, oil and grease and TKN. Due to the short test duration, no cost data were developed for the activated carbon system.

Dual Media Filtration Results

Performance Data--

Dual media filtration tests were also performed on the nitrified effluent. Runs were conducted at filtration rates of 122, 204, 326 and 407 liters/min/m^2 (3, 5, 8 and 10 gpm/ft^2). During the study, influent suspended solids averaged 190 mg/l. Removal efficiency ranged from 17 to 64 percent.

During one of the dual media filtration runs, samples of influent, effluent, and backwash were collected for metals analysis. The metals analysis data are summarized in Table 10.

TABLE 10. DUAL MEDIA FILTRATION HEAVY METALS REMOVAL

Sample point	Heavy metal concentration, mg/l						
	Cu	Cd	Pb	Zn	As	Se	Sb
Influent	0.06	0.03	<0.1	2.4	<0.0005	0.56	<0.05
Effluent	0.05	0.03	<0.1	0.9	<0.0005	0.42	<0.05
Backwash	0.86	0.13	0.20	30.0	0.0005	2.33	<0.05

As seen in Table 10 copper and selenium were removed in trace amounts, while a significant amount of zinc was removed by the filter. There were no measurable removals of arsenic, antimony or cadmium.

Cost Data--

The vendor prepared cost data for the dual media filtration system are shown in Table 11. Cost estimates were prepared for flows of 379 m^3/day (0.1 mgd), 1,892 m^3/day (0.5 mgd) and 3,785 m^3/day (1.0 mgd).

TABLE 11. DUAL MEDIA FILTRATION COST SUMMARY

Size, m^3/day (mgd)	379 (0.1)	1,892 (0.5)	3,785 (1.0)
Capital costs, (\$)[a]	114,000	237,000	387,000
Annualized costs, (\$/yr)[a,b]	34,000	50,000	73,000
Annualized costs, \$/3,785 ℓ (\$/1,000 gal.)[a, b]	0.93	0.27	0.20
Land requirement, m^2 (ft^2)	90 (950)	100 (1,100)	140 (1,500)

a. All prices are effective as of December, 1980.

b. Annualized cost = operating cost + capital amortization (20 yr life; 7.125%).

The cost estimates are based on a filtration design rate of 244.4 liters/min/m^2 (6.0 gpm/ft^2) and a suspended solids removal efficiency of 55 percent. The design backwash rate is 855 liters/min/m^2 (21 gpm/ft^2) for a 10 minute duration.

The capital cost for a 1,892 m^3/day (0.5 mgd) system would be $237,000. The annualized cost is estimated at $0.27/3.785 liters ($0.27/1,000 gal.). Space requirements for this design are estimated at 100 m^2 (1,100 ft^2).

Two Stage Activated Sludge System Performance and Cost Summary

The two stage activated sludge treatment train was successful in reducing influent concentrations of coke plant pollutant parameters. Because of upsets in the coke plant pretreatment system during the pilot study period, high dilution of the waste stream was necessary. Significant reductions in pollutant concentrations were still achieved, however. For example after dilution, an ammonia concentration of 160 mg/l resulted, which would be comparable to a well operating ammonia still effluent. Through nitrification, ammonia was reduced to an average of 15 mg/l, a 91 percent reduction. However, due to the large quantity of dilution water required, it was not possible to determine the effect of other contaminants on the nitrification process.

The treatment train consisted of: (1) diluting the raw coke plant wastewater; (2) blending the mixture for pH control and phosphorus addition; (3) removing carbonaceous material in the first stage activated sludge system followed by clarification of the wastewater with polymer addition; (4) further wastewater dilution in a ratio of 3.5 parts service water to 1 part first stage activated sludge effluent; (5) addition of sodium carbonate, powdered activated carbon and pH control chemicals to the second stage mixed liquor; and (6) oxidation of ammonia in the second stage activated sludge unit to nitrate, followed by clarification with polymer addition. Late in the study, filtration of the second stage activated sludge effluent was investigated.

Performance Summary--

The performance data summary from the pilot activated sludge investigation is shown in Table 12.

It is obvious from Table 12 that the first stage activated sludge system was effective in removing carbonaceous material as measured by phenol and BOD. There was also some thiocyanate and sulfide removal. Through the second stage activated sludge unit, ammonia, thiocyanate, suspended solids, oil and grease and phenol were reduced by significant amounts. Interferences developed in the analytical procedures which made the measurement of effluent sulfide and total cyanide impossible.

TABLE 12. PERFORMANCE SUMMARY TABLE

Parameter	Coke plant feedwater[a],	AS-I effluent	Diluted AS-II feedwater[c,d]	AS-II effluent[d]
pH	7.55[b]	7.10[b]	7.70[b]	7.30[b]
Suspended solids, mg/l	69	127	45	55
Sulfide, mg/l	31	5	0.9	I
Phenol, mg/l	655	8.3	0.3	<0.1[e]
Oil & grease, mg/l	17	10	3	9
Ammonia, mg/l	772	720	160	15
Cyanide-T, mg/l	9.9	9.2	5.0	I
Thiocyanate, mg/l	450	323	65	<1
BOD, mg/l	1,800	215	-	-

a. Diluted in the ratio of 3 parts coke plant wastewater to 1 part service water.
b. Typical value.
c. Diluted AS-I effluent in the ratio of 1 part AS-I effluent to 3.5 parts service water.
d. During period of active nitrification (January 9 - January 28, 1980).
e. Excludes data point 1/21/80 (0.6 mg/l).
I: Denotes interference.

Cost Summary--

A summary of the costs associated with activated sludge treatment of the coke plant wastewater is shown in Tables 13 and 14. Designs are presented for flows of 379 m^3/day (0.1 mgd), 1,892 m^3/day (0.50 mgd) and 3,785 m^3/day (1.0 mgd). Table 13 presents the cost information for system designs assuming treatment of the coke plant wastewater as received during the study. The Table 14 designs assume that the feedwater is effluent from a well-operating ammonia still.

From Table 13, the waste as received would have a capital cost of $5,863,000 for a 1,892 m^3/day (0.5 mgd) design. The corresponding annualized cost would be $18.68/3,785 liters ($18.68/1,000 gal.). Land requirements are estimated at 6,600 m^2 (71,100 ft^2). In comparison, if the coke plant were to provide better pretreatment of the wastewater to reduce ammonia levels, the capital cost for a similar design flow would be only $3,665,000. Similarly, the annualized costs and land requirements would be reduced to $8.33/3,785 liters ($8.33/1,000 gal.) and 3,250 m^2 (35,400 ft^2) respectively. The obvious savings should provide sufficient monetary incentive for the coke plant to better operate their existing pretreatment system.

Priority Pollutant Discussion

Priority pollutant analyses were performed on the coke plant wastewater treated in the experimental two stage biological system and the pilot activated carbon adsorption unit. A total of 13 samples plus appropriate blanks were collected during February, 1980.

Metals--

The pilot treatment system was not effective in reducing influent zinc concentrations. The mean influent concentration of zinc was 443 µg/l and the effluent 400 µg/l. Zinc was found in high concentration (630 µg/l) in the raw water added to the treatment system as makeup water. Zinc concentrations in the carbon column effluent were reduced threefold from the activated sludge effluent levels. Selenium was reduced by 77% from a mean influent concentration of 1,370 µg/l to a final effluent concentration of 320 µg/l. The selenium concentration in the carbon column effluent decreased approximately 25 percent from the activated sludge effluent level. Arsenic was reduced from a mean influent concentration of 360 µg/l to an effluent concentration of 99 µg/l. Influent silver concentration was 18 µg/l compared to an effluent level of 12 µg/l. All the other priority pollutant metals were removed to concentration levels close to or below detection limits.

Volatile Organics--

The treatment process was effective in removing all volatile organic priority pollutants. Toluene was reduced from a mean influent concentration of 607 µg/l to a final effluent of less than 10 µg/l (on two of the three sampling dates). While the activated sludge system influent contained 6,100 µg/l to 9,800 µg/l of benzene, no benzene was detected in the

TABLE 13. ACTIVATED SLUDGE COST SUMMARY - WASTE AS RECEIVED

Size, m^3/day (mgd)	379 (0.1)	1,892 (0.50)	3,785 (1.0)
Capital costs, ($) [a]			
AS-I	$ 838,000	$2,654,000	$ 4,373,000
AS-II	989,000	3,209,000	5,784,000
TOTAL	$1,827,000	$5,863,000	$10,157,000
Annualized costs ($/yr)[a,b]			
AS-I	$ 233,000	$ 823,000	$ 1,495,000
AS-II	590,000	2,587,000	5,063,000
TOTAL	$ 823,000	$3,410,000	$ 6,558,000
Annualized costs $/3,785 ℓ ($/1,000 gal.) [a,b]			
AS-I	$ 6.38	$ 4.51	$ 4.10
AS-II	16.16	14.17	13.87
TOTAL	$22.54	$18.68	$17.97
Land Requirements, m^2 (ft^2)			
AS-I	440 (4,700)	1,200 (12,900)	3,600 (38,400)
AS-II	1,210 (13,000)	5,400 (58,200)	9,100 (97,500)
TOTAL	1,650 (17,700)	6,600 (71,100)	12,700 (135,900)

a. All prices are effective as of December, 1980.

b. Annualized cost = operating cost + capital amortization (20 yr life, 7.125%).

TABLE 14. ACTIVATED SLUDGE COST SUMMARY
WELL OPERATING AMMONIA STILL FEEDWATER ASSUMED

Size, m^3/day (mgd)	379 (0.1)	1,892 (0.50)	3,785 (1.0)
Capital costs, ($) [a]			
AS-I	$ 792,000	$2,427,000	$3,965,000
AS-II	518,000	1,238,000	2,222,000
TOTAL	$1,310,000	$3,665,000	$6,187,000
Annualized costs,($/yr) [a,b]			
AS-I	$ 156,000	$ 559,000	$1,040,000
AS-II	211,000	751,000	1,425,000
TOTAL	$ 367,000	$1,310,000	$2,465,000
Annualized costs, $/3,785 ℓ ($/1,000 gal.) [a,b]			
AS-I	$ 6.05	$4.21	$3.81
AS-II	5.78	4.12	3.90
TOTAL	$11.83	$8.33	$7.71
Land requirements, m^2 (ft^2)			
AS-I	240 (2,600)	1,700 (18,700)	3,200 (34,600)
AS-II	170 (1,800)	1,550 (16,700)	2,750 (29,600)
TOTAL	410 (4,400)	3,250 (35,400)	5,950 (64,200)

a. All prices are effective as of December, 1980.
b. Annualized cost = operating cost + capital amortization (20 yr. life; 7.125%).

activated sludge or the final carbon column effluents. All other volatile organic compounds in the activated sludge and carbon column effluents were generally below the detection limits.

Base/Neutral Extractable Organics--

The treatment process was effective in reducing all monitored base/neutral extractable organics. The concentration of all nine of the base/neutral compounds found in the influent was reduced through the treatment process to less than 10 µg/l in the activated sludge final effluent.

Acid Extractable Organics--

All acid extractable organic compounds monitored were effectively removed by the treatment system. For example, the system influent contained a mean concentration of 62 mg/l phenol. Phenol was reduced by the treatment process to less than the detection limit in the final effluent. All other acid extractable organic compounds in the final effluent were below the detection limit.

DISCUSSION

The pilot program indicated that two stage activated sludge treatment of the coke plant wastewater was capable of significantly reducing influent concentrations of both priority and non-priority pollutant BAT parameters. However, a significant amount of dilution water was added to the pilot system to produce a wastewater similar in ammonia concentration to a well-operating ammonia still. Because of the problems experienced with the coke plant's ammonia still pretreatment system during the study, it is unknown if similar results would have been achieved if the research had been performed directly on the still effluent.

The cost summaries shown in Tables 13 and 14 indicate that a significant savings would result if better operation of the existing free and fixed ammonia still pretreatment system were provided by the coke plant prior to activated sludge application. For example, in a 1,892 m^3/day (0.5 mgd) design, capital costs could be reduced by $2,198,000 or 60 percent by more efficient wastewater pretreatment. Similarly, an annualized cost reduction of $10.35/3,785 liters treated ($10.35/1,000 gal. treated) would be realized. Land requirements would also be dramatically reduced. Efficiency improvement in the coke plant wastewater pretreatment system would appear to be the most logical, economical first step toward meeting the 1984 effluent guidelines.

SECTION 3

DESCRIPTION OF THE RESEARCH SITE INVESTIGATED

BY-PRODUCT COKE PLANT

General

The site chosen for this study was the Wheeling-Pittsburgh Steel Corporation's Steubenvillle East coke plant, located on the Ohio River at Follansbee, West Virginia. Coke production during the pilot study averaged 4,372 metric tons per day (4,816 tons/day) while the average wastewater flow was 1,885 cubic meters per day (0.499 mgd). The corresponding average water application rate (liters of water per metric ton of coke produced) during the study was 432 l/kkg (104 gal./ton). Daily plant production data and wastewater flow information gathered during this study is shown in Appendix A-1 and is plotted in Figure 6. A photograph of the field pilot plant setup at the coke plant is shown in Figure 7.

Wastewater Treatment System

There are two primary process water streams associated with the by-product cokemaking operations at the plant investigated:

1. Benzol plant effluent
2. Ammonia still excess liquor

A diagram of the wastewater sources and the present coke plant treatment system is shown in Figure 8.

As illustrated in Figure 8, strong ammonia liquor blowdown is stripped in the ammonia still by steam and caustic soda. Due to the steam and lime injection, the volume of this stream increases and the temperature rises to ~ 94°C (200°F). Wastewater from the benzol plant is blended with the excess ammonia liquor from the ammonia still for dilution and cooling purposes. This mixture, of which about 25 percent is excess ammonia liquor, thereby is cooled to 44-67°C (112-152°F). After passing through the equalization tank and the cooling tower, the wastewater has lost sufficient heat to make it amenable to microorganism degradation. However, downstream from the cooling tower, coal yard runoff and dilution service water are added to the wastewater stream as a plant practice.

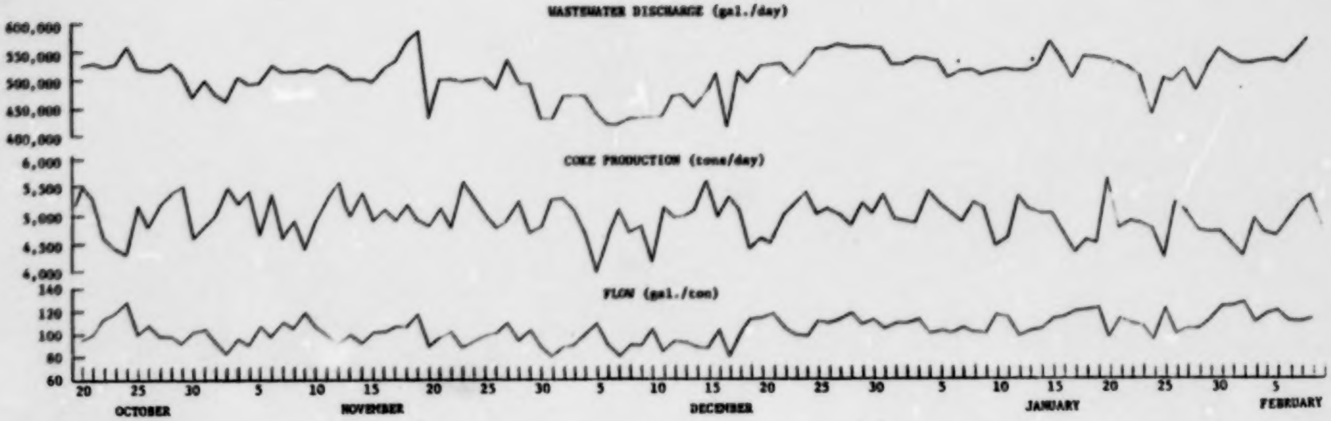


Figure 6. Wastewater discharge, coke production and flow versus date - Biological test site.

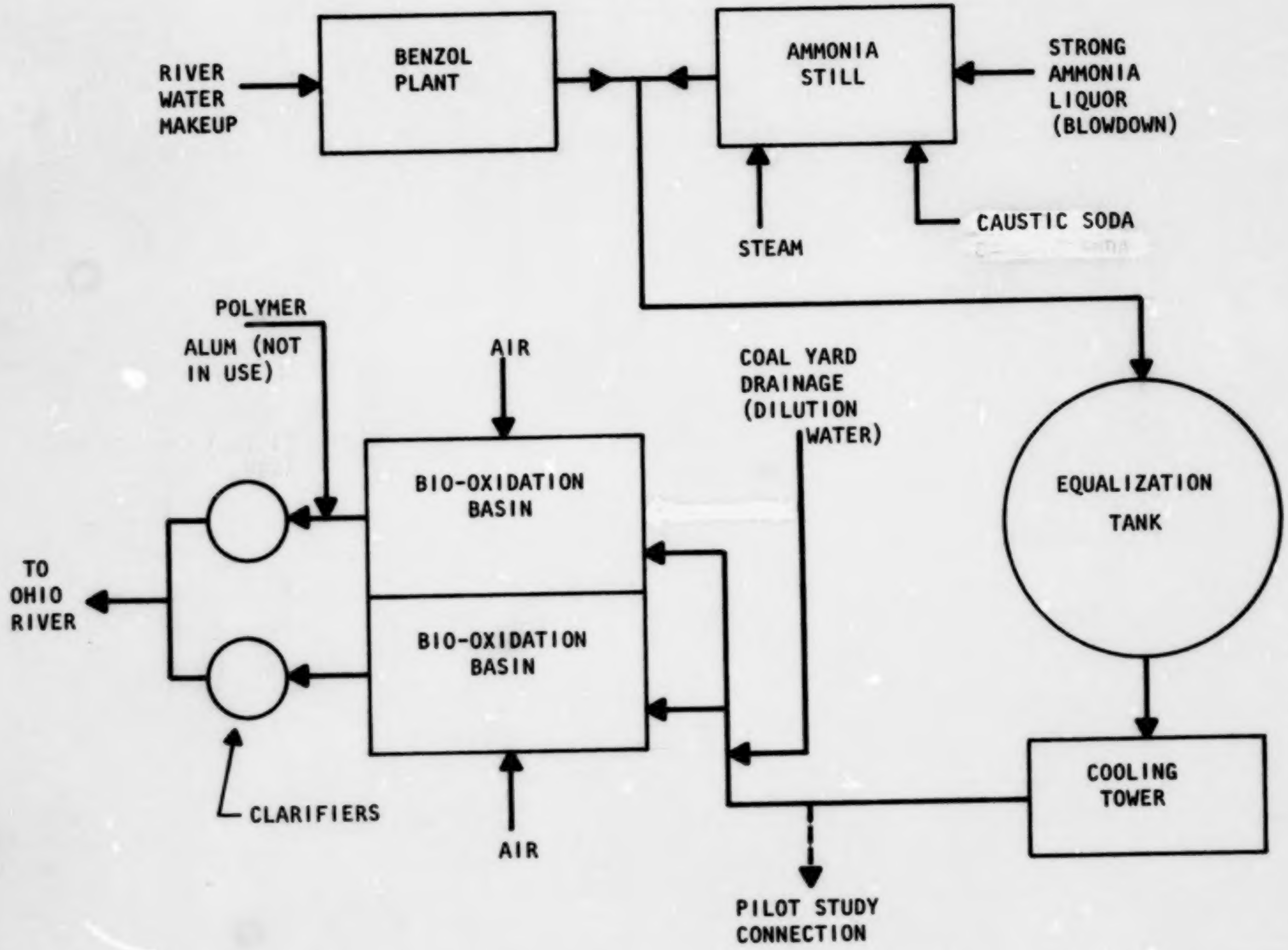

RIVER WATER MAKEUP
BENZOL PLANT
AMMONIA STILL
STRONG AMMONIA LIQUOR (BLOWDOWN)
STEAM
CAUSTIC SODA
POLYMER
ALUM (NOT IN USE)
AIR
COAL YARD DRAINAGE (DILUTION WATER)
BIO-OXIDATION BASIN
BIO-OXIDATION BASIN
EQUALIZATION TANK
TO OHIO RIVER
CLARIFIERS
AIR
COOLING TOWER
PILOT STUDY CONNECTION

Figure 8. Wastewater sources and wastewater treatment system at the by-product cokemaking plant.

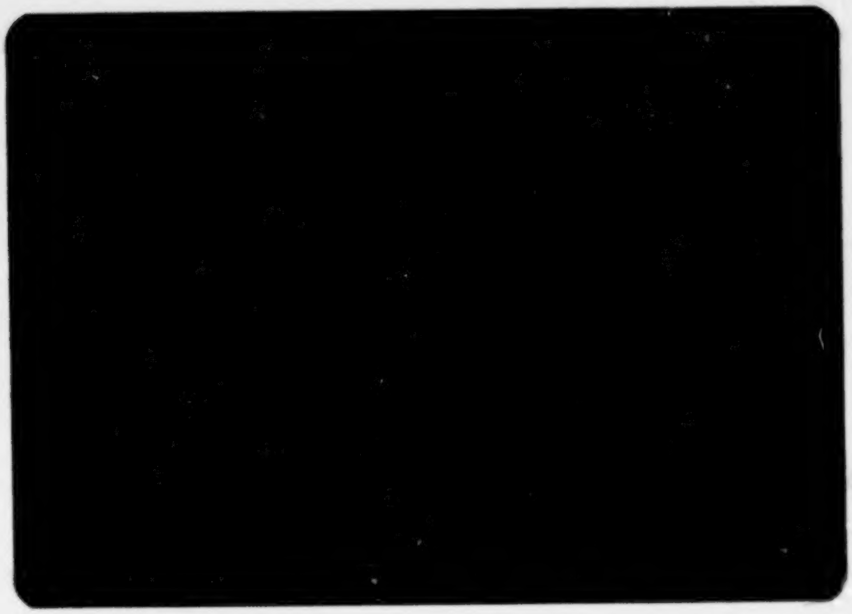


Figure 7. Field set-up - by-product cokemaking biological test site.

The wastewater is first treated by aeration in a bio-oxidation cell (concrete tank) and then flows through clarifiers where the suspended solids settle out. There is provision for alum and polymer addition, but the chemical feed system is currently not in use. Clarifier effluent is discharged to the Ohio River. The two cells of the bio-oxidation system were operated in parallel as a single stage activated sludge unit.

Best available technology (BAT) economically achievable limitations for this type of plant (Subpart D - By-Product Coke Making Plant) are indicated in Table 15.

As shown in this list, effluent limitations are based on kilograms (pounds) of pollutant discharged per 1,000 kg (1,000 lb) of coke produced.

Duration of the Study

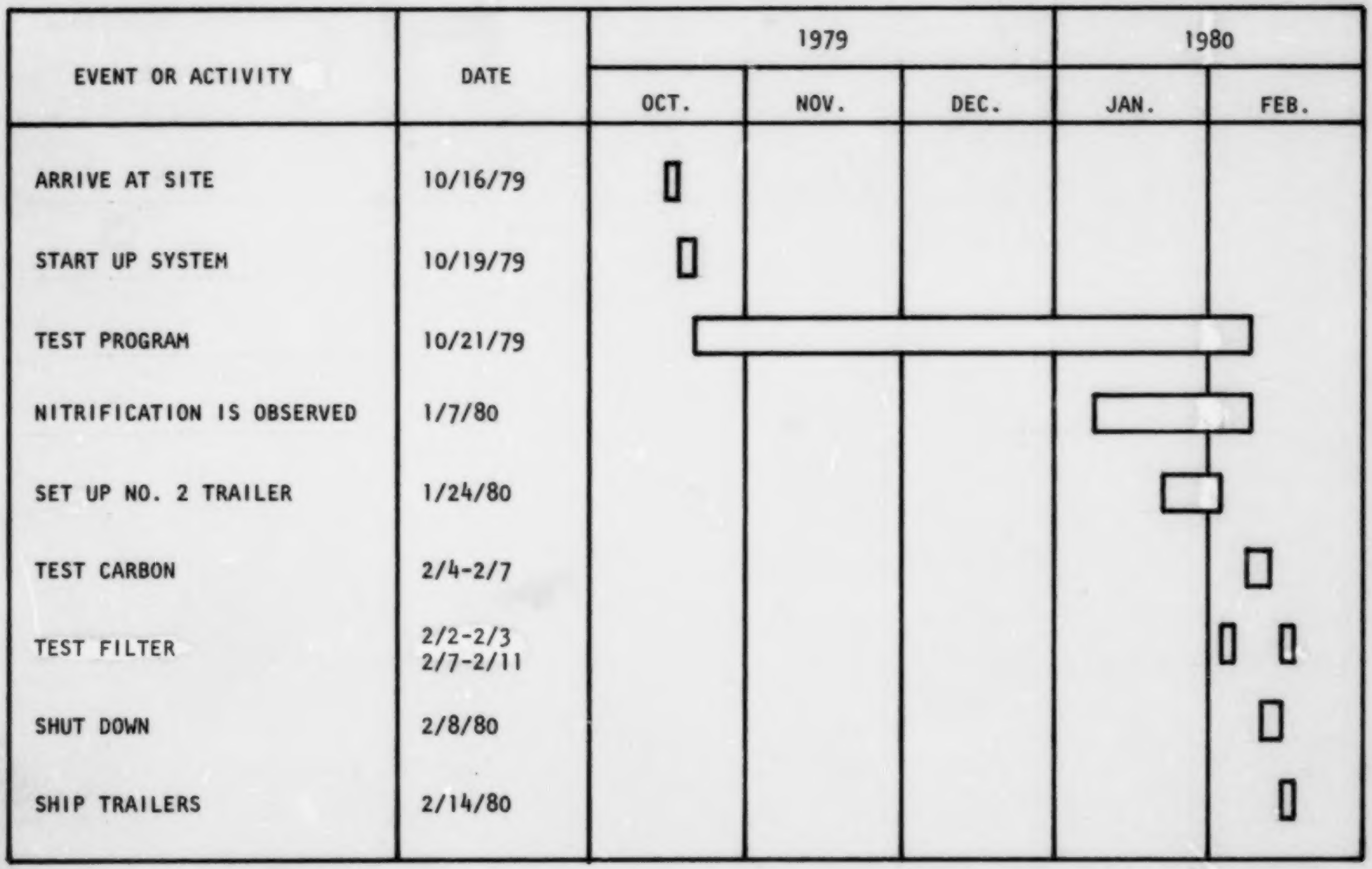
The period of time spent on-site was approximately 18 weeks, including time for startup, treatment train operation, and shutdown of the pilot plant systems. The mobile treatment units arrived at the test site on October 16, 1979. Startup began on October 18, 1979, and testing began on October 22, 1979. Operation continued 24 hours per day, 7 days per week, including the holidays, until shutdown on February 15, 1980. The field testing schedule is shown in Figure 9.

It should be noted that the wastewater temperature decreased considerably

TABLE 15. BAT - EFFLUENT LIMITATIONS GUIDELINES BY-PRODUCT COKE MAKING BIOLOGICAL TREATMENT SYSTEMS (1)

Critical parameters	Limitations kg/kkg (lb/1000 lb)	mg/l
Cyanide, Total	0.000909	1
Phenolic Compounds	0.000045	0.05
Ammonia, as N	0.0136	15
Oil & Grease	0.00909	10
Suspended Solids	0.0363	40
Sulfides	0.000273	0.3
Thiocyanates	0.000454	0.5
pH	-	6-9

during the latter half of the study, due to the on-set of winter. During the colder weather, the temperature control equipment on the trailer was used to heat the dilution water to maintain a mixed liquor temperature of 35°C (95°F). The temperature control system was required due to the large surface area to volume ratio in the pilot system. The full-scale biological plant currently in operation uses a cooling tower on the wastewater influent. The tower operates throughout the year. It is believed that the full-scale designs developed in Section 6 of this report would not require a temperature control system to maintain a suitable environment for biological growth. Rather, adjustments could be made, if needed, in the quantity of flow passed through the cooling tower to maintain an optimum wastewater temperature.



EVENT OR ACTIVITY	DATE
ARRIVE AT SITE	10/16/79
START UP SYSTEM	10/19/79
TEST PROGRAM	10/21/79
NITRIFICATION IS OBSERVED	1/7/80
SET UP NO. 2 TRAILER	1/24/80
TEST CARBON	2/4-2/7
TEST FILTER	2/2-2/3 2/7-2/11
SHUT DOWN	2/8/80
SHIP TRAILERS	2/14/80

Figure 9. By-product coke plant pilot plant investigation schedule.

SECTION 4

WASTE STREAM CHARACTERIZATION

The coke by-product wastewater at the biological testing site consisted primarily of wastewater from the benzol plant and excess ammonia liquor from the ammonia still. A grab sample of the wastewater was obtained by Rexnord on September 13, 1979. The only other available information concerning the composition of the coke by-product wastewater was from the priority pollutant sampling that was done by NUS Corporation in September 1977. The NUS samples were taken just before the cooling tower and the Rexnord sample was taken immediately after the cooling tower (Refer to Section 3 for a plant schematic). The cooling tower was a non-contact enclosed type, and therefore, no significant change in waste composition would be expected through the cooling tower.

The average composition of the samples obtained by NUS and the composition of the grab sample obtained by Rexnord are listed in Table 16. Also included in Table 16 are the composition of the activated sludge plant effluent as measured by NUS in 1977.

Comparison of the analyses of the raw waste samples shows that the Rexnord grab sample obtained in 1979 contained significantly higher concentrations of ammonia, thiocyanate, phenol, and organic carbon than the composite samples obtained by NUS in 1977. According to plant officials, this was due to improper operation of the ammonia stills during the testing period. The dissolved solids concentrations were similar and the sulfide content of the 1979 sample was much lower. The results of the NUS analyses of the plant effluent shows that the activated sludge plant was very effective in removing phenols, sulfides, organic carbon, and thiocyanates. These good removals may be due to the low activated sludge plant influent concentrations. The raw wastewater is diluted with coal yard drainage prior to activated sludge treatment. Based on the composition of the 1979 grab sample, it appears that dilution prior to biological treatment would be necessary. Typical intake water analyses from the NUS survey are shown in Appendix A-2.

TABLE 16. COMPOSITION OF COKE BY-PRODUCT WASTEWATER SAMPLES

Sample	Raw Waste	Raw Waste	Plant Effluent
Sample Date	9/13/79	Sept. 1977	Sept. 1977
Sampled By	Rexnord	NUS	NUS
Sample Type	Grab	Average of 3 composites	Average of 3 composites
Ammonia N, mg/l	990	148	127
TKN, mg/l	1,200	-	-
Cyanide, mg/l	40	12.1	2.4
Thiocyanate, mg/l	640	336	0.8
Sulfide, mg/l	0.6	325	0.2
Phenol, mg/l	940	520	0.05
TOC, mg/l	1,700	787	74
BOD, mg/l	3,110	-	-
Dissolved solids, mg/l	10,310	10,610	3,657
Chloride, mg/l	6,600	4,943	1,693
pH	8.8	9.6	7.6
Alkalinity, mg/l	1,168	-	-

SECTION 5

TREATMENT TECHNOLOGIES INVESTIGATED

INTRODUCTION

This report section describes the equipment used to perform the pilot scale program at the coke plant. The primary treatment equipment consisted of a two-stage biological oxidation unit, an activated carbon column and a dual media filtration system. Discussion in this section regarding pollutant treatment, sampling and analysis is limited to conventional and non-conventional parameters only. A separate discussion on toxic pollutant results is presented in Section 7.

The sampling procedure, data collection points, and the major analytical parameters tested for during the study are presented in Table 17. It should be noted that all of the indicated parameters were not run on all of the samples. Typically, the raw waste was sampled daily throughout the study period. The various intermediate streams (activated sludge units 1 and 2 mixed liquor, return sludge, clarifier effluent, etc.) were sampled over selected periods. A more detailed description of the sampling and analytical schedule for each process is presented in the Section 6 Data Evaluation Section.

BIOLOGICAL OXIDATION

Introduction

The use of biological oxidation for the removal of phenols and organic carbon from coke by-product wastewaters is accepted practice. Most such facilities, however, provide little or no removal of ammonia nitrogen. Attempts at ammonia removal by biological nitrification have met mixed success. Some of the problems have been attributed to the sensitivity of nitrifying organisms to toxic substances and to sudden changes in waste concentration and temperature. Another problem with nitrification units is the difficulty in maintaining the desired concentration of nitrifiers in the system because of the poor settling characteristics of the mixed liquor.

The addition of powdered activated carbon (PAC) has been used successfully to improve the performance of activated sludge plants treating chemical and refinery wastes containing toxic substances (3, 4). In addition, the

TABLE 17. TYPICAL ANALYTICAL SCHEDULE FOR BIOLOGICAL PILOT STUDY

Parameter	Sample type	Sample points[1]
pH	composite	All points
Suspended solids	composite	1,2,3,4,5,6,7
Volatile suspended solids	composite	1,2,4,5,6,7
Ammonia	composite	1,3,6
Kjeldahl nitrogen	composite	1,3,6
Nitrate-nitrite	composite	1,3,6
Total phosphate	composite	1,3,6
Cyanide	composite	1,3,6
Thiocyanate	composite	1,3,6
Phenol	composite	1,3,6
Alkalinity	composite	1,3,6
Oil and grease	grab	1,3
TOC	composite	1,1S,3,3S,6,6S
BOD	composite	1,3,6

1. See Figure 10 for location of sampling points. S after a number indicates that the analysis was performed on the soluble portion of the sample.

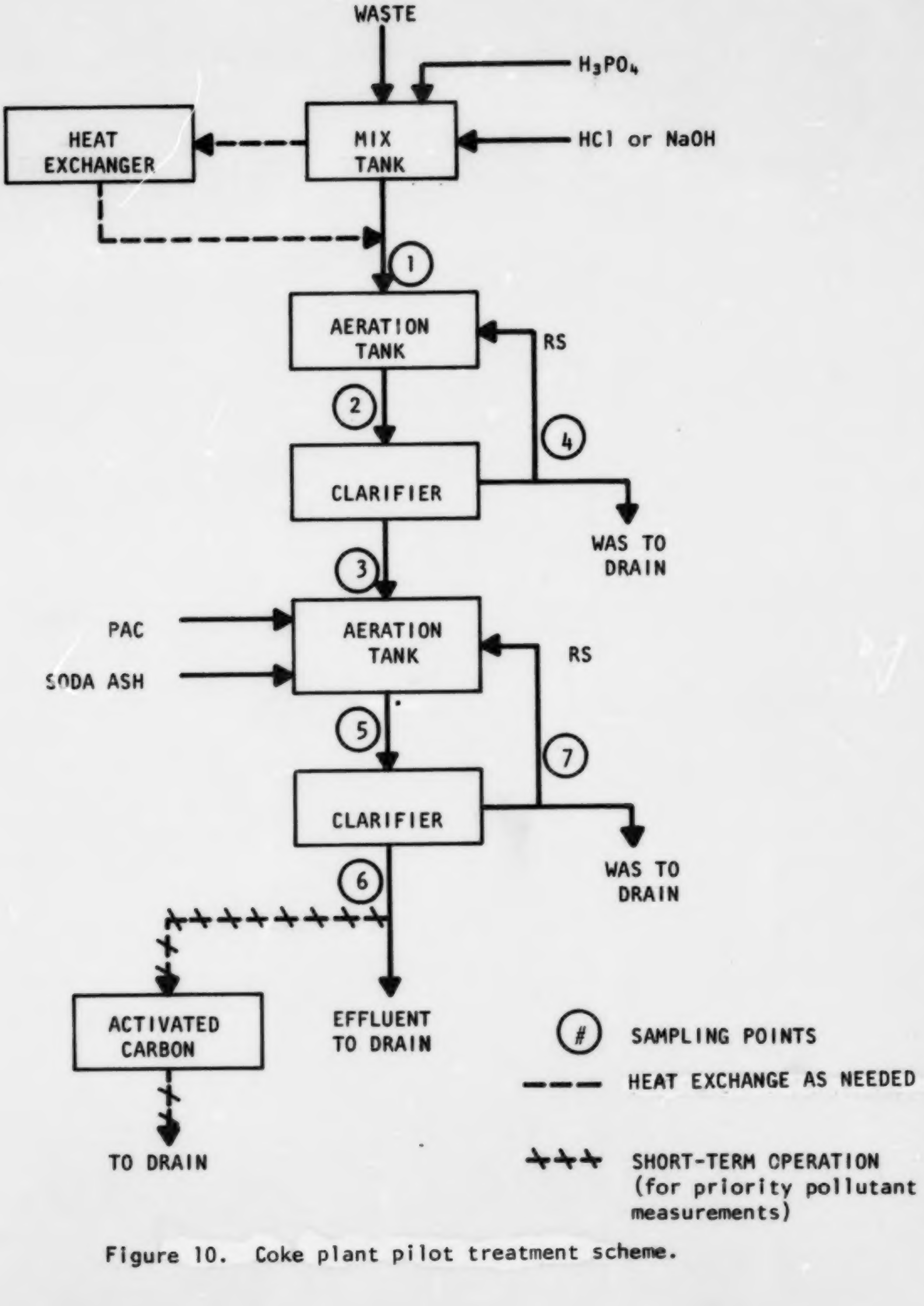

WASTE
H3PO4
HEAT EXCHANGER
MIX TANK
HCl or NaOH
1
AERATION TANK
RS
2
4
CLARIFIER
WAS TO DRAIN
3
PAC
SODA ASH
AERATION TANK
RS
5
7
CLARIFIER
WAS TO DRAIN
6
ACTIVATED CARBON
EFFLUENT TO DRAIN
TO DRAIN
# SAMPLING POINTS
HEAT EXCHANGE AS NEEDED
SHORT-TERM OPERATION (for priority pollutant measurements)

Figure 10. Coke plant pilot treatment scheme.

PAC apparently reduces the impact of sudden changes in waste strength and improves the settleability of the mixed liquor.

Pilot plant tests were performed on coke by-product wastewater at the Wheeling-Pittsburgh Steel Company plant at Follansbee, West Virginia to evaluate a two stage activated sludge treatment system. The first stage was operated as a conventional activated sludge plant for the purpose of removing organics, phenols, and some toxic materials. The second stage was a nitrification activated sludge plant with powdered activated carbon added to the mixed liquor to improve stability and performance. The pilot plants were operated with the objective of determining the effluent quality that can be achieved under several loading and operating conditions and obtaining data needed for estimating the cost of treatment.

Pilot System Description

As illustrated in Figure 10, the bio-oxidation system consisted of an influent mixing tank, a heat exchanger, and two identical activated sludge systems. The heat exchanger was required to warm the influent wastewater during cold weather. Nutrient chemicals were added to the mix tank along with acid or caustic soda for automatic pH control.

Each activated sludge system had an aeration tank, air blower, clarifier, and appurtant pumps, valves, and instrumentation as shown schematically in Figure 11. The self-priming pump on the first activated sludge system pumped chemically treated wastewater from the mix tank through the heat exchanger into the aeration tank. Aerated wastewater flowed by gravity to the clarifier, where overflow (effluent) was collected and pumped to the second system. Sludge was continuously recycled by pump from each clarifier cone bottom to each aeration tank inlet. A portion of the return sludge was wasted. Positive displacement lobe type compressors continuously pumped air into headers located in each aeration tank to mix and aerate the mixed liquor. The aeration tanks were divided into 3 equal compartments. Chemicals were added by duplex chemical feeders from two mix tanks on each system.

Except for different chemicals used, both activated sludge systems were essentially identical. The following table lists the rates and capacities of the activated sludge pilot equipment.

Feed Pump	19 l/min (5 gpm)
Recycle Pump	19 l/min (5 gpm)
Aeration Tank	
Total Volume	5,942 liters (1,570 gal.)
Detention Time @ 19 l/m (5 gpm)	5.2 hr

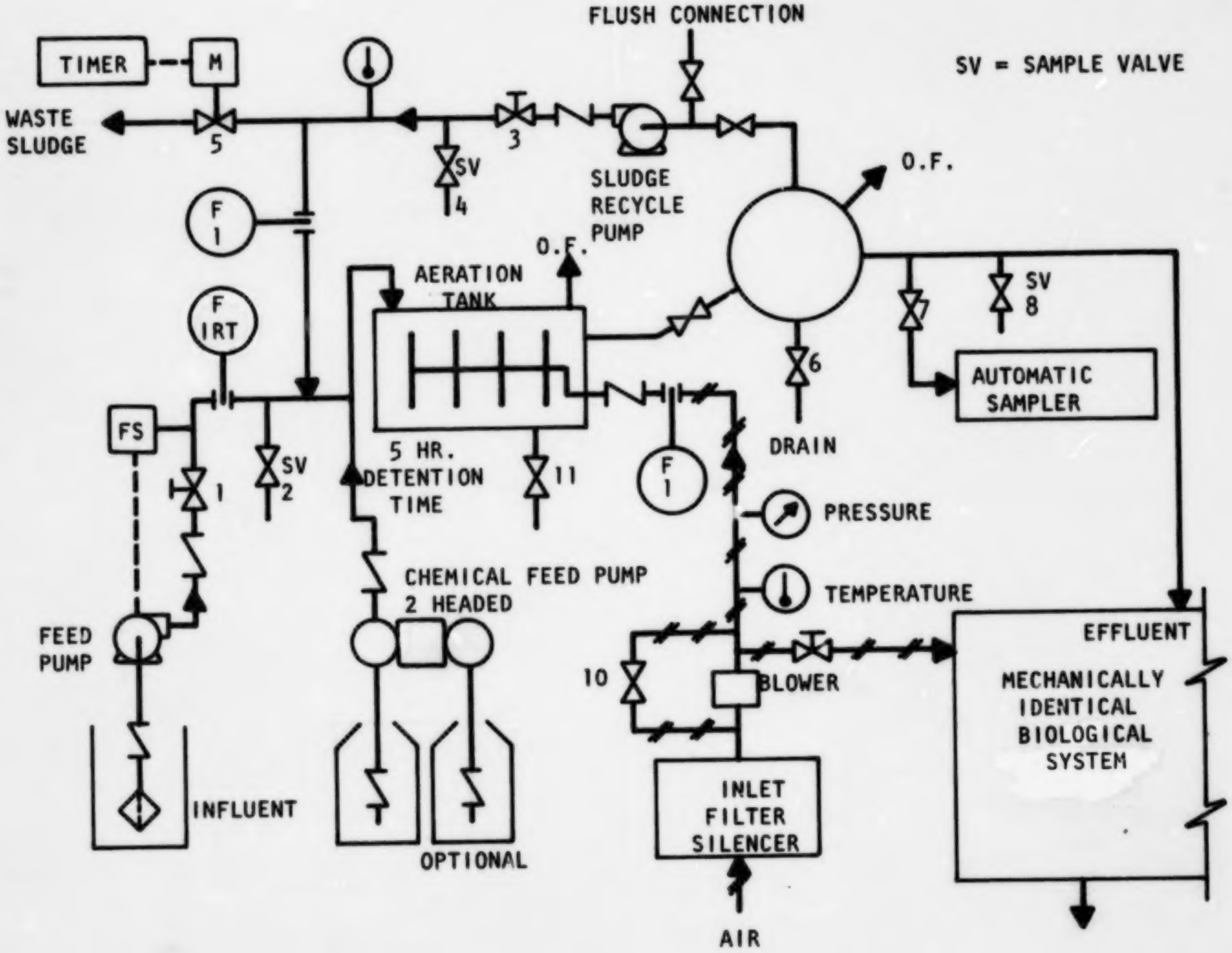

FLUSH CONNECTION
SV = SAMPLE VALVE
TIMER
M
WASTE SLUDGE
5
3
SV 4
SLUDGE RECYCLE PUMP
O.F.
F 1
O.F.
AERATION TANK
F IRT
SV 8
7
6
AUTOMATIC SAMPLER
FS
SV 2
1
5 HR. DETENTION TIME
11
F 1
DRAIN
PRESSURE
TEMPERATURE
CHEMICAL FEED PUMP 2 HEADED
FEED PUMP
EFFLUENT
10
BLOWER
MECHANICALLY IDENTICAL BIOLOGICAL SYSTEM
INFLUENT
INLET FILTER SILENCER
OPTIONAL
AIR

Figure 11. Biological treatment system - first stage.

Clarifier	
Diameter	130 cm (4'-3")
Overflow Rate, @ 19 l/min (5 gpm)	174 l/d/m^2 (500 gpd/ft)
Volume	2,839 liters (750 gal.)
Detention Time @ 19 l/m (5 gpm)	2.5 hr
Chemical Pump	0-30 l/hr (0-8 gph each head)
Blower (Factory Ratio)	0.84 m^3/min @ 3,515 kg/m^2 (30 scfm @ 5 psi)
(Factory Pulleys)	1.68 m^3/min @ 3,515 kg/m^2 (60 scfm @ 5 psi)

Experimental Program

The pilot plants were operated with the objectives of determining the effluent quality that could be achieved while operating a two stage activated sludge system under typical coke plant conditions. Operating data were also necessary for estimating the cost of treatment. In addition, the use of powdered activated carbon (PAC) was evaluated as a means for improving performance of coke plant second stage activated sludge system. Due to the short test duration, it was not possible to investigate second stage performance both with and without activated carbon, and therefore, PAC was used throughout the study.

The first stage of the activated sludge system was operated as a conventional activated sludge plant for the purpose of removing organics, phenols, and some toxic substances. The second stage was operated as a nitrification activated sludge plant with powdered activated carbon added to the mixed liquor to improve stability and performance.

In order to facilitate the start-up of the biological pilot plant, mixed liquor from the aeration tank of the full-scale treatment plant was used to fill the first stage aeration tank. The first stage was then fed raw wastewater which was in a ratio of 3 parts coke plant wastewater to 1 part service water. Dilution is commonly practiced in the biological treatment of coke plant wastewater. The particular ratio selected was based on the dilution ratio used in the full-scale plant. In this manner, a high quality effluent from the first stage was obtained in a very short time and enhanced the start-up of the pilot plant second stage (nitrification).

The following discussion describes the equipment operation in detail:

1. Mix Tank - The raw waste from the outlet of the cooling tower was piped to a small mixing tank where dilution water was added, the desired amount of phosphoric acid was added and the pH adjusted with either hydrochloric acid or (when necessary) sodium hydroxide. A schematic of the mix tank is shown in Figure 12.

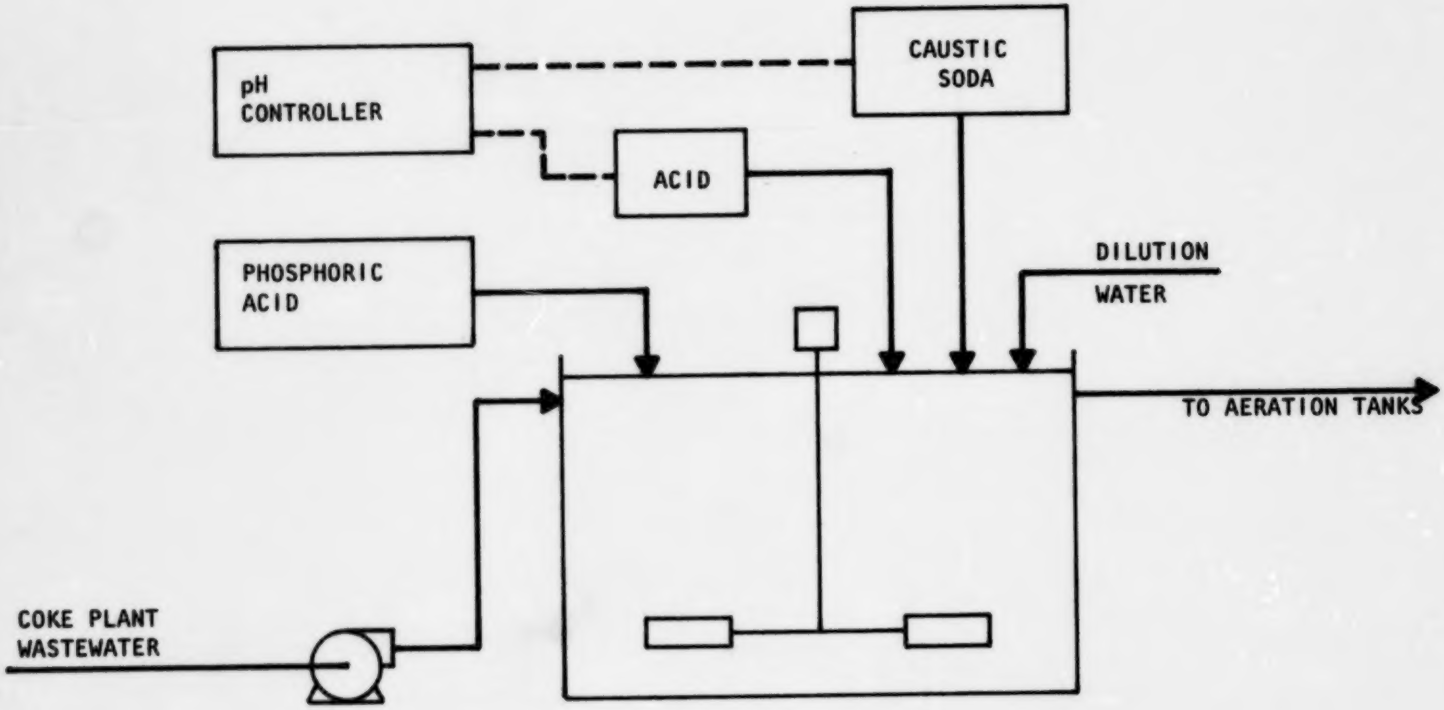

pH
CONTROLLER
CAUSTIC
SODA
ACID
PHOSPHORIC
ACID
DILUTION
WATER
TO AERATION TANKS
COKE PLANT
WASTEWATER

Figure 12. Flow diagram of coke plant bio-oxidation mix tank system.

a. Phosphate addition - Initially, phosphoric acid was added to the mix tank at a rate such that the influent to the first stage aeration tank contained 10.6 mg/l P. After start-up, the soluble ortho-phosphate concentration in the mixed liquor leaving the second stage aeration tank was adjusted to maintain a small (about 0.2 to 0.5 mg/l) ortho-phosphate residual.

b. Dilution water addition - Tap water was added to reduce the raw waste concentration to approximate the operation of the full-scale treatment facility. Prior to entering the first stage activated sludge unit, three parts of raw wastewater were diluted with one part of tap water. Dilution was also required prior to the second stage activated sludge unit (nitrification). Here the wastewater was diluted to maintain a constant feed ammonia strength (∿ 200 mg/l) to the system.

c. pH adjustment - A Balsbaugh pH controller was used to control the pH of the first stage activated sludge unit to approximately 7.0. The pH adjustment was made in the neutralization tank upstream of the first stage. A small mechanical mixer ensured thorough mixing of the contents and provided improved automatic pH control. The flow from the mix tank was pumped to the activated sludge system through the pilot unit heat exchanger. The temperature of the waste entering the aeration tank was typically maintained between 32 and 35°C. Flow measurements were made at the inlet to the activated sludge system.

2. Activated Sludge System - The activated sludge system consisted of two units. The first unit (AS-1) was designed for carbonaceous BOD reduction, and was operated as a conventional activated sludge plant.

The second unit (AS-11) was used for the nitrification of ammonia. Powdered activated carbon was used in this unit to reduce the effect of toxic materials on the nitrifiers and improve the settling characteristics of the activated sludge. During the startup of the second stage unit, the wastewater was dosed with a mutant bacteria specially manufactured to enhance the growth of nitrifying organisms.

An intensive testing and monitoring effort was performed after the units reached a steady-state condition. The purpose of this work was to characterize the performance of the units for the removal of pollutants and the tolerance to potentially inhibiting substances. During this period, selected tests were performed as needed to estimate the physical size and performance of a full-scale system. The intensive testing period was performed over a ten day period.

The samples taken and the analyses performed were listed previously in Table 17.

In addition, the following special tests were performed during the test period.

Settling rate and SVI test - Performed daily on grab samples of mixed liquor from AS-1 and AS-11. Samples of the mixed liquors used in these tests were submitted to the laboratory for suspended solids analysis.

Waste activated sludge thickening tests - Thickening tests were performed on grab samples of waste sludge from AS-1 and AS-11 during the intensive testing period. Gravity thickening was investigated. Samples for suspended solids analysis were sent to the laboratory for analysis.

Oxygen uptake rate tests - A series of oxygen uptake rate tests were performed on samples of return sludge from AS-1 and AS-11 during the intensive testing period. Each series of tests included 3 or 4 F/M ratios (ratio of diluted waste to return sludge solids). This information was used in equipment sizing for the full-scale system. Samples of soluble TOC and suspended solids from these tests were sent to the laboratory for analysis.

The tasks listed in Table 18 were used as a guide by the plant operators for the work to be performed on-site. The operators used their judgment and the advice of the office back-up man at the home office to determine the relative priority of the tasks. Office and home phone numbers of back-up people at the home office were provided for emergency.

3. Priority Pollutant Sampling - At selected times during the pilot tests, samples were taken for analysis of priority pollutants. The analytical work was performed by Southern Research of Birmingham, Alabama. Sampling took place during an intensive testing period and was performed by an outside contractor.

   During the intensive testing period for each bio-treatment system, the effluent was passed through a carbon column to determine the effect of carbon adsorption on the removal of priority pollutants. The bio-treatment effluents were passed through the carbon columns for a one week period. Because of the short time available for carbon column effluent testing, carbon break-through data were not obtained. At routine intervals during the operation of the carbon column on the biologically treated effluent, samples for priority pollutant analyses were obtained.

4. Data Records - Data sheets were used to record and organize the pilot plant operating data. A new sheet was used for each day's operation.

TABLE 18. DAILY OPERATION ROUTINE

Work Activity	Initial Start-up Period	Interim Period	Intensive Monitoring Period
1. Measure and record flows, temperature and chemicals used. Adjust if necessary.	daily	daily	daily
2. Measure and record dissolved oxygen concentration in each activated sludge aeration cell. Adjust if necessary.	daily	occasionally	daily
3. Check chemical solution stock and prepare new solutions as needed.	daily	daily	daily
4. Obtain samples and perform field analysis:			
a. Run MLSS and MLVSS on samples from points 2 and 3.	daily	daily	daily
b. Measure pH on samples from points 1,3,5, and 6.	daily	daily	daily
c. Measure ortho-phosphate at points 1 and 6.	daily	alternate days	daily
d. Measure alkalinity at points 1, 3 and 6.	daily	occasionally	daily
e. Measure ammonia at points 1, 3 and 6.	occasionally	alternate days	daily
f. Measure thiocyanate at points 3 and 6.	occasionally	occasionally	daily
5. Make process changes indicated by results of field analysis	daily	daily	daily
6. Prepare samples for shipment to laboratory. Clean samplers and check desicant.	daily	daily	daily

(continued)

Refer to Figure 10 for Sample Locations

TABLE 18. (continued)

Work Activity	Initial Start-up Period	Interim Period	Intensive Monitoring Period
7. Perform normal equipment maintenance as specified in equipment manual.	daily	daily	daily
8. Perform settling, thickening and/or oxygen uptake tests.	none	occasionally	as listed in text
9. Obtain data on tons of production and gallons of wastewater for the previous day	daily	daily	daily
10. Maintain accurate and complete records of pilot equipment operation, field analysis and sample collection and shipment	daily	daily	daily

Field analytical data recorded during the study were entered in the bound Field Laboratory Book. All calculations were included in the book.

All other pertinent information was entered in the pilot unit log book. In particular, items such as breakdowns, unusual operating conditions, etc. were entered.

ACTIVATED CARBON

Introduction

Activated carbon adsorption was investigated as a post-treatment technology to the two stage biological oxidation system. The study was performed during an intensive testing period when the biological units were operating at steady-state. The primary purpose of the investigation was to determine the efficiency of carbon to remove toxic pollutants from the coke plant wastewater. The carbon columns used to conduct the study were located in the EPA physical-chemical mobile wastewater treatment van.

Because of the short duration of testing, carbon column breakthrough data were not obtained. Samples for priority pollutant analyses were collected by an outside contractor and sent to Southern Research of Birmingham, Alabama for analysis.

Description of Pilot Equipment

The activated carbon column system was equipped with four 35.6 cm (14 in.) diameter columns. However, for this study, only one of the four columns was filled with activated carbon to a depth of 152 cm (5 ft). A schematic diagram of the carbon adsorption system is shown in Figure 13.

Each carbon column is provided with media strainers in the top and bottom of the vessel. The strainers trap the media within the vessel and serve to distribute the flow. The piping is designed to allow operation in either the upflow or downflow mode. For this test, the downflow mode was chosen. Sample taps are located on 30.5 cm (12 in.) centers along the column, but samples were taken only from the bottom 3 taps which correspond to the media depth. The self-priming activated carbon system pump fed the influent wastewater into the top of the first column. Effluent from the column flowed through the bottom strainers to the drain. The unused columns and piping were thus isolated during this test.

The following table lists the capacities and rates of the carbon adsorption system.

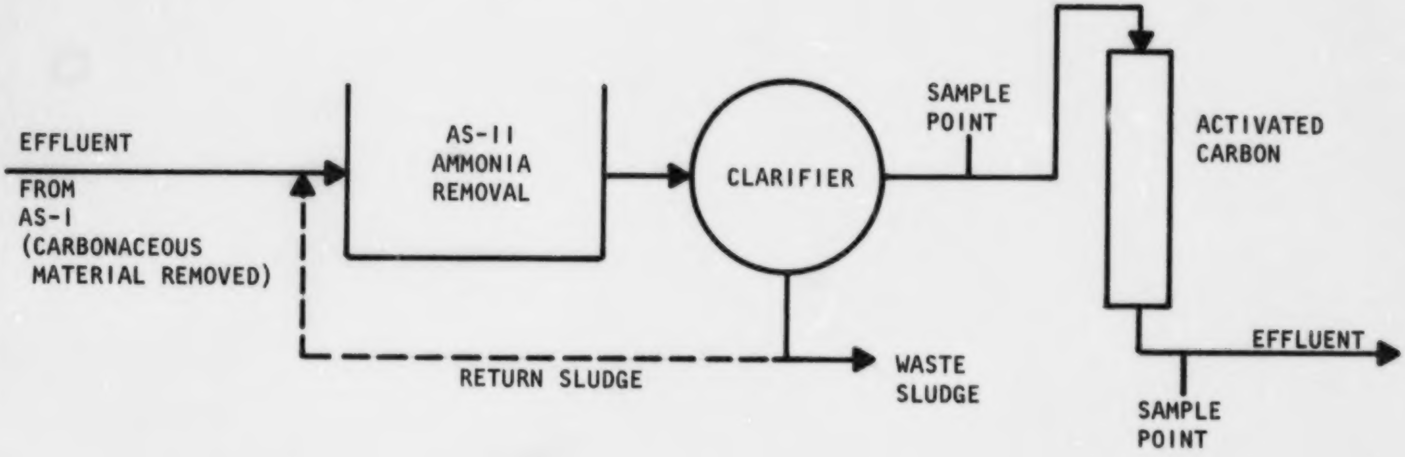

EFFLUENT
FROM
AS-I
(CARBONACEOUS
MATERIAL REMOVED)
AS-II
AMMONIA
REMOVAL
CLARIFIER
SAMPLE
POINT
ACTIVATED
CARBON
EFFLUENT
SAMPLE
POINT
RETURN SLUDGE
WASTE
SLUDGE

Figure 13. Schematic of activated carbon system used during study.

Feed Pump	19 l/min @ 38,665 kg/m^2 (5 gpm @ 55 psi)
Backwash Pump	38 l/min @ 21,090 kg/m^2 (10 gpm @ 30 psi)
Column Area	0.09 m^2 (1.0 ft^2)
Filter Rate	1.74 l/min/m^2 (5 gpm/ft^2)
Backwash Tank	1,892 liters (500 gal.)
Air Scour	0.14 m^3/min @ 3,515 kg/m^2 (5 scfm @ 5 psi)
Bed Expansion	30%
Minimum Size Granules	0.5 mm (0.020 in.)
Column Design Maximum Pressure	42,180 kg/m^2 (60 psi)
Carbon Depth	152 cm (5 ft).
Carbon Volume	0.14 m^3 each column (5 ft^3 each column)

Experimental Program

The activated carbon testing procedure consisted of passing a measured volume of wastewater through a given bed depth of carbon. Sampling and analysis were an important part of the program. Composite samples of the influent and effluent were taken on a regular basis. Analysis for conventional pollutants included: pH, TOC, suspended solids, volatile suspended solids, color, TKN, ammonia, phosphorus, cyanide - total and amenable, oil and grease, BOD, and thiocyanate. In addition, samples were collected for toxic pollutants.

The mode of operation selected for the pilot carbon study was downflow. Direction of wastewater flow does not affect overall carbon adsorptive performance. Sampling access could be obtained at any of the provided taps. When the waste reached the bottom of the column, it passed through two neva-clog screens. The influent and effluent screens were so constructed as to prevent the escape of carbon without impeding liquid flow. For backwashing, clean water entered at the bottom of the column, flowed upward and loosened the entrapped solids for discharge at the top of the column.

To provide an accurate record of daily operation, a standard sheet was used to record information during the study.

DUAL MEDIA FILTRATION

Introduction

To provide background data for coke plant effluent suspended solids removal by filtration, a small pilot size dual media filter was brought to the test site. The removal of heavy metals and oil and grease was also studied. The dual media filter was used to test contaminant removal from the second stage bio-oxidation system (nitrified effluent). The filtration tests were the last tests performed at this site.

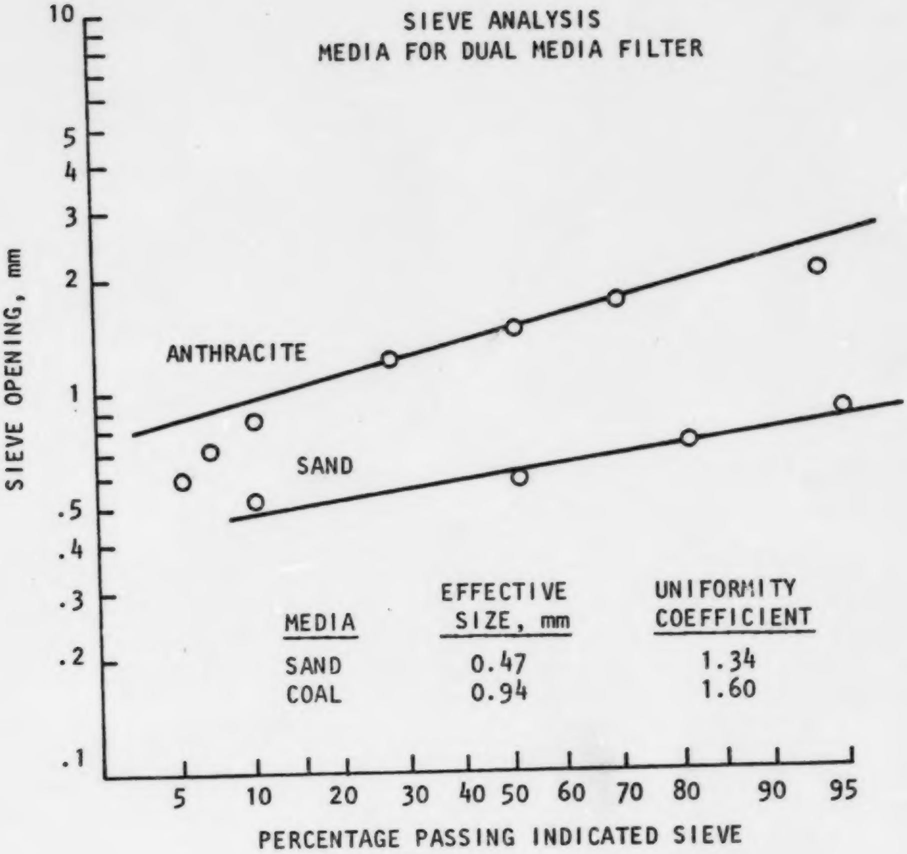
Description of Pilot Equipment

The pilot filter was constructed of clear plexiglas plastic tube with a 7.6 cm (3 in.) diameter. The filter area was 0.0046 m^2 (0.05 ft^2). Filter media consisted of 43 cm (17 in.) of filter sand covered with 50.8 cm (20 in.) of anthracite coal. Antracite coal was purchased from The Reading Anthracite Coal Company, Pottsville, Pennsylvania. The material specified was to have a 0.80 to 1.20 millimeter effective size and maximum uniformity coefficient of 1.65. Filter sand was purchased from the Red Flint Sand Company, Eau Claire, Wisconsin. The material specified was to have a 0.45 to 0.55 millimeter effective size and a uniformity coefficient of 1.4. Before loading the media, representative samples were taken by the "cone and quartering" method for sieve analysis. The results are presented in Figure 14. The anthracite media used in the pilot program had an effective size of 0.94 millimeter with a 1.60 uniformity coefficient, while the sand media effective size was 0.47 millimeter with 1.34 uniformity coefficient.

The carbon column self-priming pump was used to pressurize the dual media filter. As shown in Figure 15, pressure gauges and sample taps were provided at the influent and effluent points, and within the media. The media gauges were placed near the top of the anthracite layer, at the anthracite/sand interface, and near the midpoint of the sand layer. Tap water was used to backwash the filter media.

Experimental Program

The objective of pilot filter tests was to obtain sufficient information for full scale design. Of vital interest were performance characteristics for determining effluent quality and design parameters to insure the most cost effective equipment size. Performance characteristics included effluent quality and the consistency of effluent during varying feed conditions. The filter surface area required to process the design flow, backwash water processing requirements, feed preconditioning requirement, power requirements, and filter cleaning requirements are examples of design parameters which can be obtained from pilot testing. Pertinent independent variables include the following:



MEDIA	EFFECTIVE SIZE, mm	UNIFORMITY COEFFICIENT
SAND	0.47	1.34
COAL	0.94	1.60

Figure 14. Filter media sieve analysis.

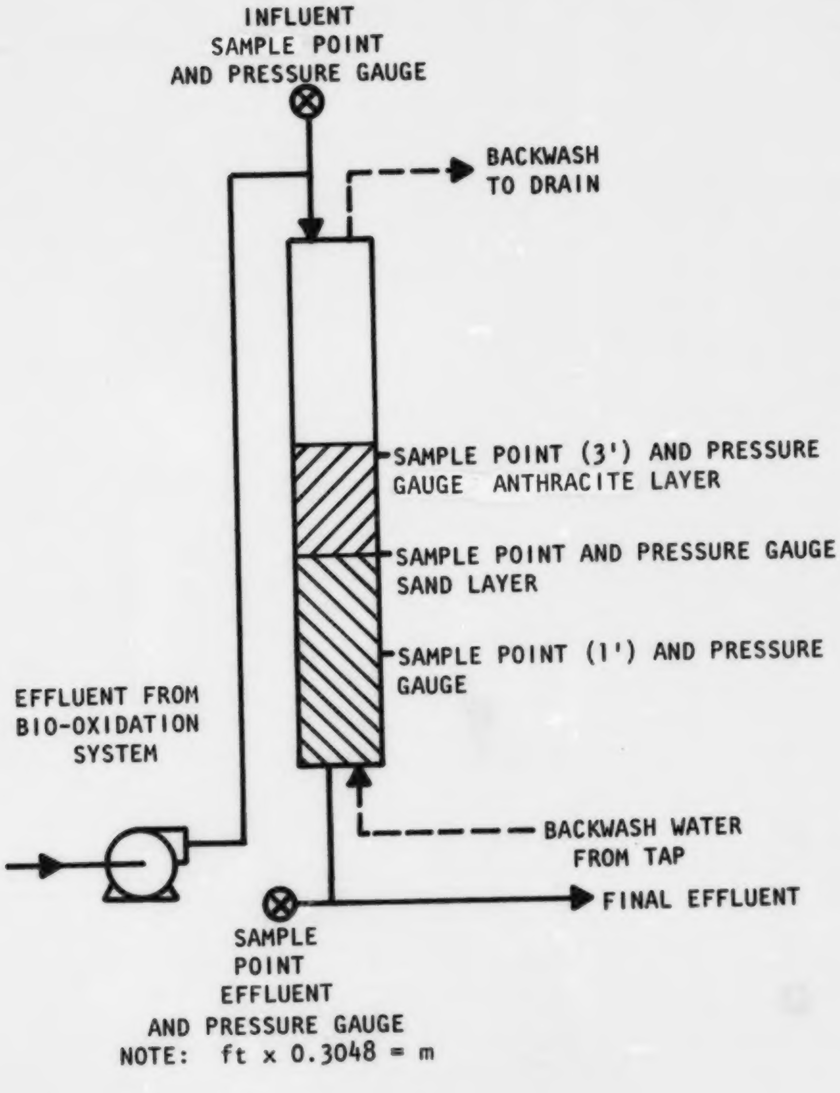

INFLUENT
SAMPLE POINT
AND PRESSURE GAUGE
BACKWASH
TO DRAIN
SAMPLE POINT (3') AND PRESSURE
GAUGE ANTHRACITE LAYER
SAMPLE POINT AND PRESSURE GAUGE
SAND LAYER
SAMPLE POINT (1') AND PRESSURE
GAUGE
EFFLUENT FROM
BIO-OXIDATION
SYSTEM
BACKWASH WATER
FROM TAP
FINAL EFFLUENT
SAMPLE
POINT
EFFLUENT
AND PRESSURE GAUGE
NOTE: ft x 0.3048 = m

Figure 15. Small dual media filter test system.

1. Filtration rate, l/min/m^2 (gpm/ft^2) filter area.
2. Feed characteristics.
3. Media selection.
4. Backwash.
5. Mode of filter operation.
6. Terminal headloss.
7. Chemical preconditioning.

Pilot testing determines how these independent variables affect the following dependent variables.

1. Measurement of production.
    a. weight of solids captured per unit filter area as a function of headloss
    b. net volume of water filtered as a function of headloss
2. Filtrate quality.
3. Cause of filtration termination.
    a. headloss
    b. solids breakthrough
4. Depth of solids penetration as indicated by a headloss profile of filter bed.
5. Effectiveness of backwash.

The data obtained during each filter run were recorded on a standard data sheet. Routine recordings included time, flow rate, headloss profiles, pH, and temperature.

Based on accepted filtration practice, the following test conditions were selected.

1. Filtration Rate: The filtration rate was varied from 122 to 407.4 l/min/m^2 (3 to 10 gpm/ft^2) throughout the study period.

2. Feed Characteristics: This variable is determined by the process preceding the filters. During the study, the dual media filter processed biologically treated effluent.

3. Filter media: On loading the media, a known quantity of sand was placed in the filter and backwashed at 611 liters/min/m^2 (15 gpm/ft^2) for 30 minutes. To remove fine material, the top 5 cm (2 in.) was removed. Next, a known amount of anthracite was placed on the sand and backwashed and 5 cm (2 in.) removed. The total media depth was then recorded.

4. Backwash: A relatively mild degree of backwashing was used throughout the study. Tap water was used to backwash the filter at 611 l/min/m^2 (15 gpm/ft^2) for about ten minutes. Each backwash was initiated manually so that a given filter run would correspond to a sampling interval.

5. Mode of Filter Operation: All tests were run downflow under pressure with constant head and constant rate.

The filter was operated continuously throughout the study. Each interval between backwashing was called a run and assigned a number. Run lengths were generally determined by the sampling schedule and were approximately 8 to 10 hours long. There were occasional runs of longer or shorter duration as the flow was set to the desired rate and initial headloss readings taken. Headloss profiles through the media were recorded at intervals through the run and at termination. Terminal headloss rather than breakthrough point typically dictated the end of a run.

Samples of feed and filtrate were composited for each run for suspended solids determination. At the termination of a run the filter was backwashed and prepared for the next run. Samples of the backwash water from one run were taken during the water wash cycle and analyzed for suspended solids.

SECTION 6

COKE PLANT DATA EVALUATION

INTRODUCTION

This section presents a summary of the findings at the coke plant research site. For each technology investigated (activated sludge-carbonaceous removal, activated sludge-nitrogen removal, filtration, and activated carbon), a general section has been provided for the process introduction, a presentation and discussion of the pilot study results, a full scale system description and full scale system costs and space requirements.

It should be noted that throughout the study, pollutant values in the coke plant wastewater remained abnormally high due to malfunction of the free and fixed ammonia stills. As a result, high dilution of the waste stream was required. Thus, an artificial operating condition existed which significantly reduced the value of the study results. Because of this, it is impossible to draw conclusions as to the ability of the pilot system to meet BAT requirements. However, substantial pollutant reductions were achieved and a concerted effort has been made to incorporate all significant findings into this report.

The field study consisted of an extensive sampling and analytical program. A summary of all analytical results obtained during the study is presented in Appendices B-1 and B-2. A separate discussion on the sampling and analytical procedures used in the study is presented below. The discussion in this section is limited to conventional and non-conventional pollutants only. Toxic pollutant sampling and analytical results are presented in Section 7.

LAB QUALITY CONTROL

Samples taken during the study periods consisted of grab and 24 hour composites. Grab samples were taken for parameters such as oil and grease. Composite samples were obtained by collecting hourly discrete samples using ISCO automatic samplers over 24 hour periods.

Throughout the sampling program, the samples were iced during collection and shipment. Pertinent preservation methods in accordance with EPA (5) were performed in the field immediately after collection. Samples for Biochemical Oxygen Demand (BOD) were split and submitted to a local laboratory. The remaining samples were packaged, iced, and sent to the Rexnord

Inc. Environmental Research Center (ERC) laboratory in Milwaukee, Wisconsin. Chances of sample mis-labeling were eliminated by use of unique sample identification numbers, a field sample log, and the ERC analytical laboratory sample log.

The ERC laboratory also maintains a quality assurance program. Table 19 presents the calculated population standard deviations for all duplicate analyses performed during the study. The duplicate analyses have been divided into concentration ranges or populations to better illustrate the change in precision with concentration.

Method accuracy was monitored throughout the study using standards and spiked samples. Method accuracy, shown as percent recovery, is listed for all major parameters in Table 19. Standard deviations and percent recoveries for all parameters are similar to the precision and accuracy limits given in Standard Methods (6), indicating that each respective procedure was in control during the test program.

All analytical procedures used during the study period were referenced from Standard Methods for the Examination of Water and Wastewater (6). The equipment used and the specific reference for all analytical procedures is presented in the following list.

1. pH - pH meter, Standard Methods, p. 460.
2. Turbidity - Hach Model 2100A turbidimeter, Standard Methods, p. 132.
3. Color - Hellige Aquatester with color comparison discs calibrated at 0, 5, 10, 20, 30, 40, 50, 60, and 70 APHA color units. Waste samples were filtered and diluted with distilled water (if necessary) to reduce color to less than 70 APHA units. When the color of the sample was clearly between that of two color discs, the values for both color discs were recorded. For example, the term 0/5 means that the color of the sample was more than 0 APHA units, but less than 5 APHA units. For plotting and calculation purposes, the value midway between the two color discs was used. Standard Methods, p. 64.
4. BOD - Standard Methods, p. 543.
5. Alkalinity - Standard Methods, p. 278.
6. Total Organic Carbon (TOC) - Beckman Model 915 TOC Analyzer, Standard Methods, p. 532.
7. Suspended and Volatile Suspended Solids - Standard Methods, p. 94 and 96.
8. Oil and Grease - Standard Methods, p. 515.
9. Total Phosphate - Standard Methods, p. 476 and 624.
10. Total Dissolved Solids - Standard Methods, p. 93.
11. Sulfate - Standard Methods, p. 496.
12. Chloride - Standard Methods, p. 304.
13. Conductivity - Standard Methods, p. 71.
14. Metals - Na, Si, K, Mn, Fe, Al, Ca, Mg - Standard Methods, p. 144.

TABLE 19. SUMMARY OF STATISTICAL EVALUATION OF LABORATORY QUALITY CONTROL ANALYSIS

Parameter	Concentration range, mg/l	Mean, mg/l	Coefficient of variation	Standard deviation	Percent recovery
SS	0-50	27	6.91	1.9 (33)[a]	
	51-100	71	4.33	3.1 (45)	
	101-250	162	3.90	6.3 (32)	
	>250	9,683	3.09	299.2 (27)	
	30-33				110.0
	236-239				101.2
	548-572				95.8
TDS	1-10,000	3,237	0.52	17.0 (1)	
	>10,000	16,028	0.20	32.2 (4)	
	54-57				105.6
	323-338				95.6
	487-502				97.0
TS	1-1,000	4,945	0.45	22.3 (4)	
	>1,000	22,231	0.53	117.7 (16)	
CN_T	0-1.0	0.63	1.13	0.01 (3)	
	>1.0	85	2.19	1.9 (4)	
	0.225-0.300				98.2
	0.65-1.00				65.0
	4.3-5.0				86.0
	88-100				91.0
NH_3	0-15	10	1.97	0.2 (9)	
	16-100	30	2.04	0.6 (7)	
	>100	503	2.64	13.3 (12)	
	0.90-1.00				90.0
	1.05-1.13				92.9
	1.70-1.80				105.8
	5.9				100.0
	12.8-13.4				95.5
	14.8-16.3				90.8
	24.0-24.6				102.5
	33.8-35.8				94.4
	437				90.0
	1,568				92.3
TKN	0-250	92	0.99	0.9 (4)	
	>250	478	1.63	7.8 (4)	
	0.89-0.90				98.9
	1.13-1.20				106.2

(continued)

TABLE 19. (continued)

Parameter	Concentration range, mg/l	Mean, mg/l	Coefficient of variation	Standard deviation	Percent recovery
TKN	3.9				100.0
	10.2-10.95				93.2
	11.0-11.7				94.0
	13.4-14.6				91.7
	16.8-17.5				96.0
	33.7-36.9				91.3
	48				100.0
NO_2/NO_3	0.55-0.58				96.0
	0.89				100.0
	1.03-1.04				100.9
	1.08-1.14				100.5
O&G	57-59				96.6
Phenol	0-100	95	4.47	4.2 (1)	
	0.13				100.0
	0.30				100.0
	0.35-0.36				97.2
S	0-30	18.75	1.95	0.4 (2)	
	41.6-43.1				103.6
TOC	138-145				99.9
Cl	35-37				101.5
	53-55				103.8
	122-123				99.2
	219				100.0
	230				100.0
	900-925				102.7
	2,310-2,333				99.0
	5,038-5,136				100.5

[a] Indicates the number of standard deviations used to calculate population standard deviation.

15. Fluoride - Standard Methods, p. 391.
16. Chemical Oxygen Demand - Standard Methods, p. 550.
17. Ammonia Nitrogen - Standard Methods, p. 407.

DESCRIPTION OF COST ANALYSIS

Delivered costs for the various principal pieces of equipment were obtained from vendor quotations and from a recent edition of the Dodge Manual for Building Construction Pricing and Scheduling (7). The total plant cost includes all of the funds required to procure equipment, land and buildings in addition to funds required to design, build and bring the facility to acceptable operation. To determine the total plant cost, the Chilton method (8) of cost estimation was utilized. The Chilton method uses a factored approach. That is, the delivered costs of the principal pieces of equipment in the system are summated and then multiplied by various factors appropriate for the type of system being built. Factors for most items have a selection range which requires the judgement of the estimator to make an appropriate selection within the range.

Similar factors were selected for all systems which permits consistent comparison between systems.

The particular Chilton factors used for the various treatment methods and a costing example is listed in Appendices C and D.

The principal items that are included in the cost estimates are the following:

1. Capital Equipment.
2. Buildings.
3. Land Acquisition and Site Preparation.
4. Utilities and Yard Services (air, water, electricity, etc.).
5. Engineering Services.
6. Contractor's Fees.
7. Contingencies.
8. Winterization.

The assumptions that were used in preparing the cost estimates are listed in Table 20.

TABLE 20. ASSUMPTIONS USED TO DEVELOP COST AND DESIGN FIGURES FOR FULL SCALE SYSTEMS

1. Electric costs: $0.03/KWH

2. Capital amortization: 7.125 percent
   20 year service life

3. Labor: $19,000/year skilled
   $16,000/year unskilled

4. The costs developed are exclusive of supervisory and laboratory expenses, personnel overhead, fringe benefits, indirects (interest on capital during construction, etc.), sludge handling and disposal, etc.

5. All prices are effective as of December 1, 1980.

RESULTS OF THE ACTIVATED SLUDGE (CARBONACEOUS REMOVAL) STUDY

Introduction

The first stage of the biological oxidation treatment system (AS-1) used to treat coke by-product wastewater was successful in removing phenol and organic carbon. This stage provided influent to the second stage of biological treatment for the purpose of ammonia removal.

General Results

Due to the nature of biological treatment, a start-up period was required to encourage the growth of organisms responsible for pollutant removal. In order to facilitate start-up, mixed liquor from the aeration tank of the full-scale treatment plant was used to fill the first stage aeration tank. The first stage was then fed raw wastewater diluted to the same concentration as the feed to the full-scale plant (3 parts coke plant wastewater to 1 part service water). This dilution is normal practice for the coke plant. In this manner, a high quality first stage effluent was obtained in a reduced time which also enhanced the start-up of the second stage.

Once start-up was achieved, it was apparent that a number of operational factors affected the unit performance. Early in the study, problems were encountered with foam build-up. On at least three occasions AS-1 aeration tanks foamed over causing a decrease in MLSS concentration and a corresponding reduction in BOD removal efficiency. Use of a commercial defoamer eliminated this problem.

Operational difficulties were also encountered due to tar buildup in feed lines and valves, causing a decrease in influent flow. This flow variation was minimized through regular cleaning and valve replacement and did not cause any major upsets in treatment efficiency.

Others factors which affected the operation of AS-I were pH, dissolved oxygen in the mixed liquor, temperature and waste strength. Each of these factors and others were maintained within a specific range of values in order to optimize treatment efficiency. Brief discussions of each of these are presented below.

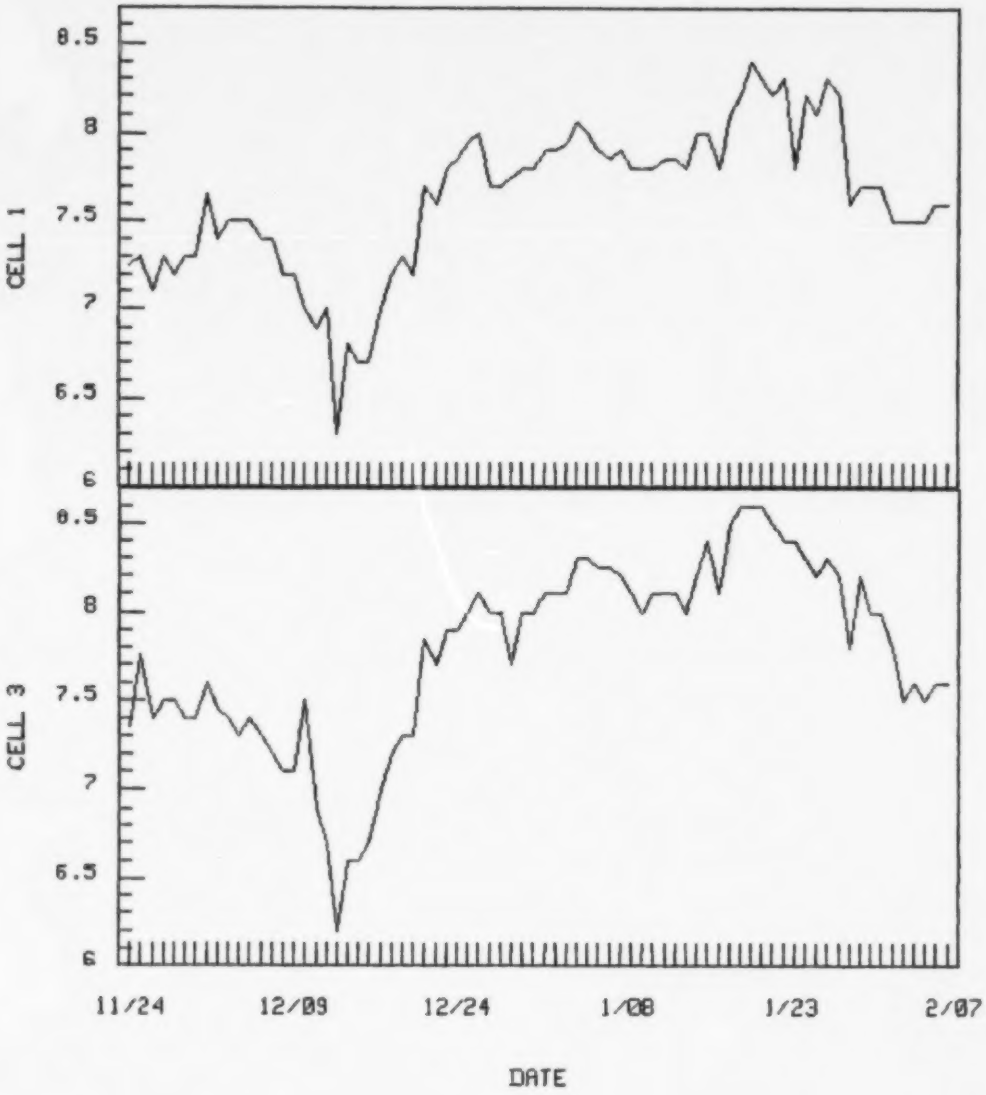
pH--
In order to maintain AS-I pH at approximately 7.0, acid addition was generally required. Only on one occasion did the coke plant influent pH drop below 7.0. The pH control was achieved through the use of an automatic controller having provision for addition of sulfuric acid or caustic soda. Shown below in Table 21 are the range and typical pH values encountered during the study for the raw undiluted waste and cell 1 and cell 3 of AS-1. Complete pH profiles can be found in Appendix Table B-3.1.

TABLE 21. AS-1 pH DATA SUMMARY

Location	pH Range	Typical pH
Raw undiluted	4.7-10.3	8.9
Cell I AS-I	6.3-8.4	7.6
Cell 3 AS-I	6.2-8.6	7.8

As can be seen the pH remained relatively constant through the cells of AS-I, although from Appendix B-3.1, it should be noted that the pH of the raw waste inexplicably went up throughout the study. A plot of cell 1 and cell 3 pH can be seen in Figure 16.

Temperature--
In order to maintain an active biomass aeration basin temperature was maintained between 25 and 35 °C. During the study this was accomplished through the use of dilution and a heat exchanger. Undiluted waste was, in most cases, above the desired temperature range although dilution generally brought the temperature into the acceptable range. Table 22 demonstrates the temperature range, mean and standard deviation for raw undiluted waste, diluted waste from the mixing tank and cell 1 and cell 3 of AS-1. Complete temperature profiles for the entire study can be found in Appendix B-3.2.



(COMPOSITE SAMPLE DATA)

Figure 16. Aeration basin pH - AS I.

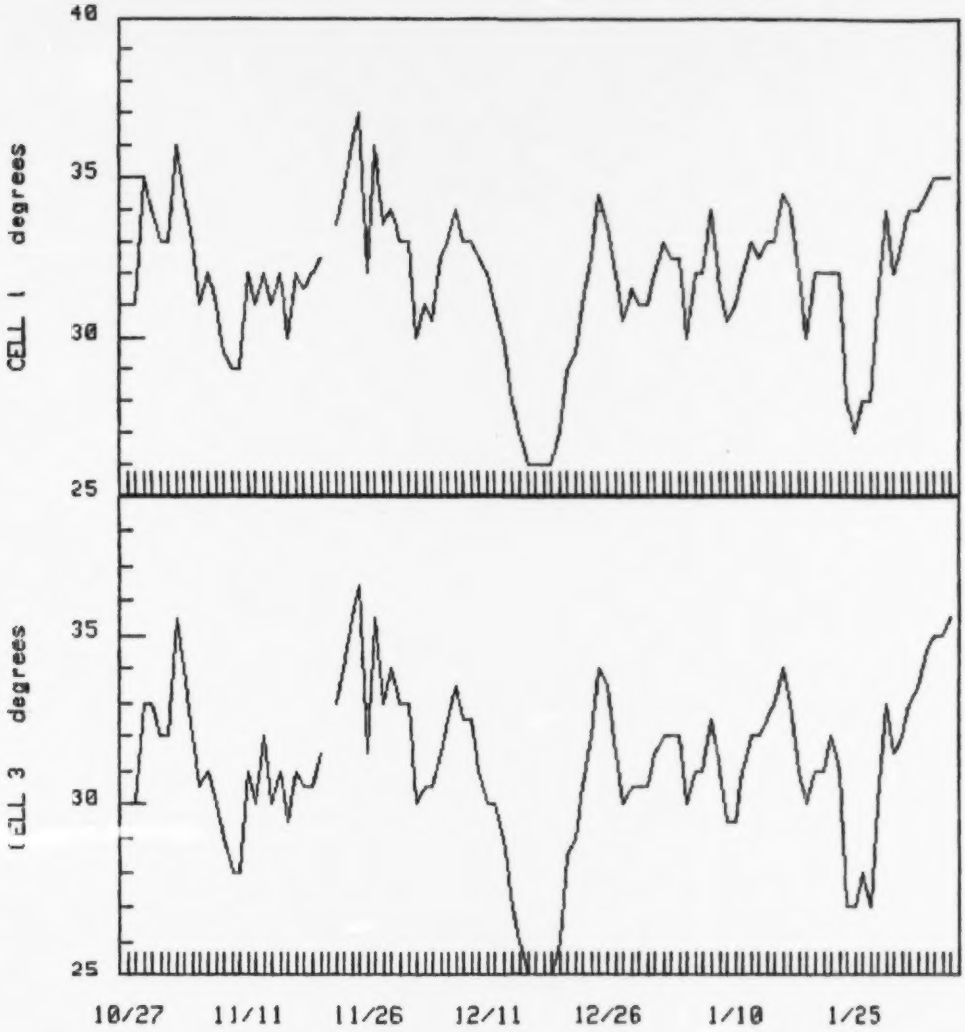
TABLE 22. TEMPERATURE DATA SUMMARY - AS-1

Location	Range, °C	Mean, °C	Std. Dev., °C
Undiluted raw	25.5-50	40.4	4.3
Mixing tank	28-42	36.2	2.7
Cell 1	26-37	31.4	3.7
Cell 3	25-36.5	31.3	2.4

As can be seen from above, the temperature generally remained quite steady through the system, and well within the desired range. Figure 17 consists of a plot showing the daily variability in cell 1 and cell 3 temperature during the project.

Dissolved Oxygen--

Maintaining adequate dissolved oxygen (DO) is another factor which is crucial for efficient biological oxidation. In this study oxygen was assimilated by the waste very efficiently because of the large amount of organic matter present. Due to this it became very difficult to provide a high DO residual in cell 1 of AS-1. Dissolved oxygen in cell 1 typically ranged from 0.5-2.0 mg/l. This DO level increased as oxygen demand decreased through cells 2 and 3. This increase in dissolved oxygen can be seen in Figure 18 and Table 23. Table 23 shows the mean and standard deviation of DO levels in cells 1, 2 and 3. A complete summary of dissolved oxygen data is provided in Appendix B-3.3.



(GRAB SAMPLE DATA)

Figure 17. Aeration basin temperature (^{o}C) - AS-1.

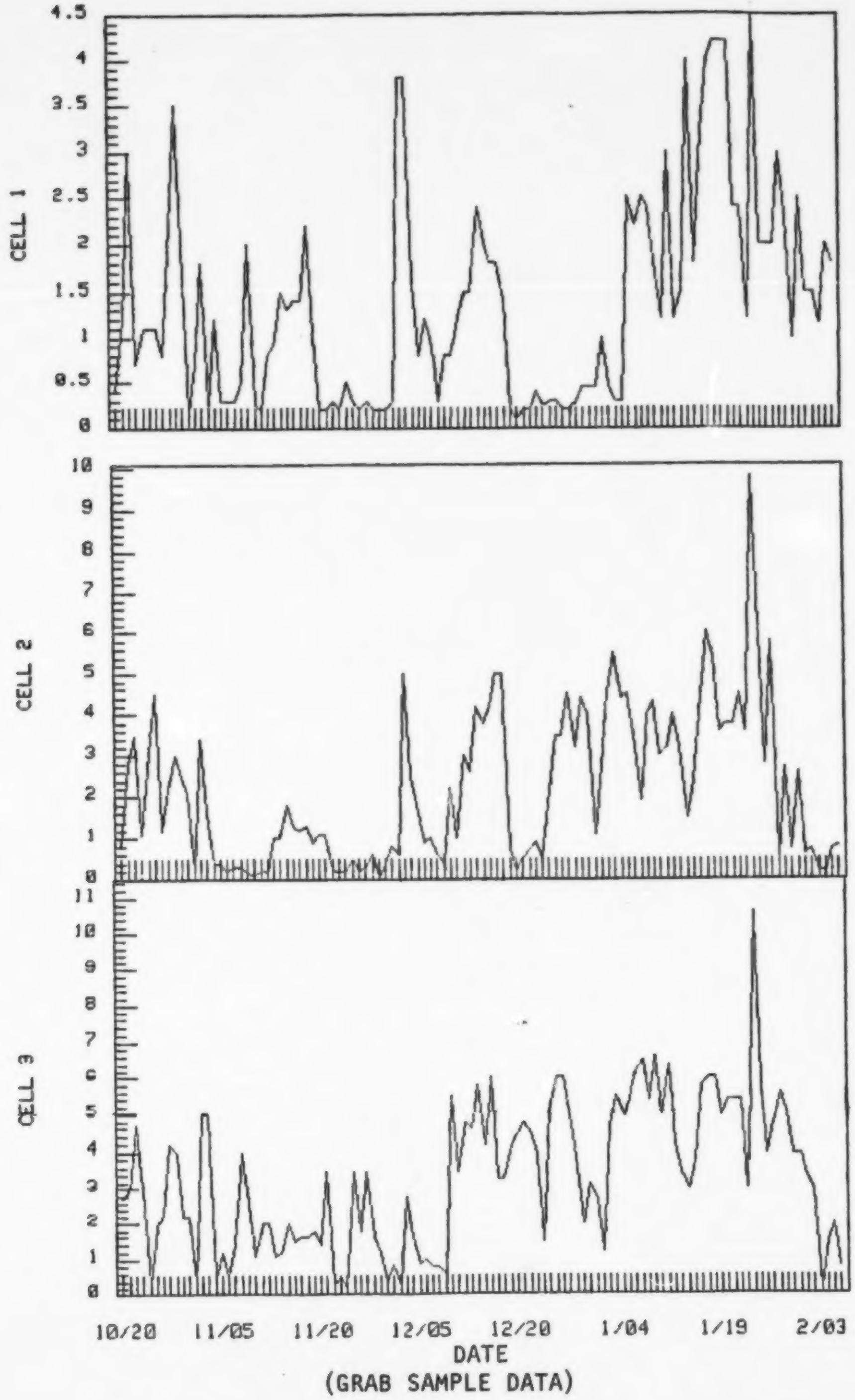


Figure 18. Aeration basin dissolved oxygen (mg/l) - AS-I.

TABLE 23. DISSOLVED OXYGEN DATA SUMMARY

Location	Mean, mg/l	Std. Dev., mg/l
Cell 1	1.4	1.2
Cell 2	2.2	1.8
Cell 3	3.3	2.0

It is apparent from the figure that, in general, the amount of dissolved oxygen available for bio-oxidation increased as the study progressed. This is due in part to a decrease in the amount of organics present in the waste.

Waste Strength--

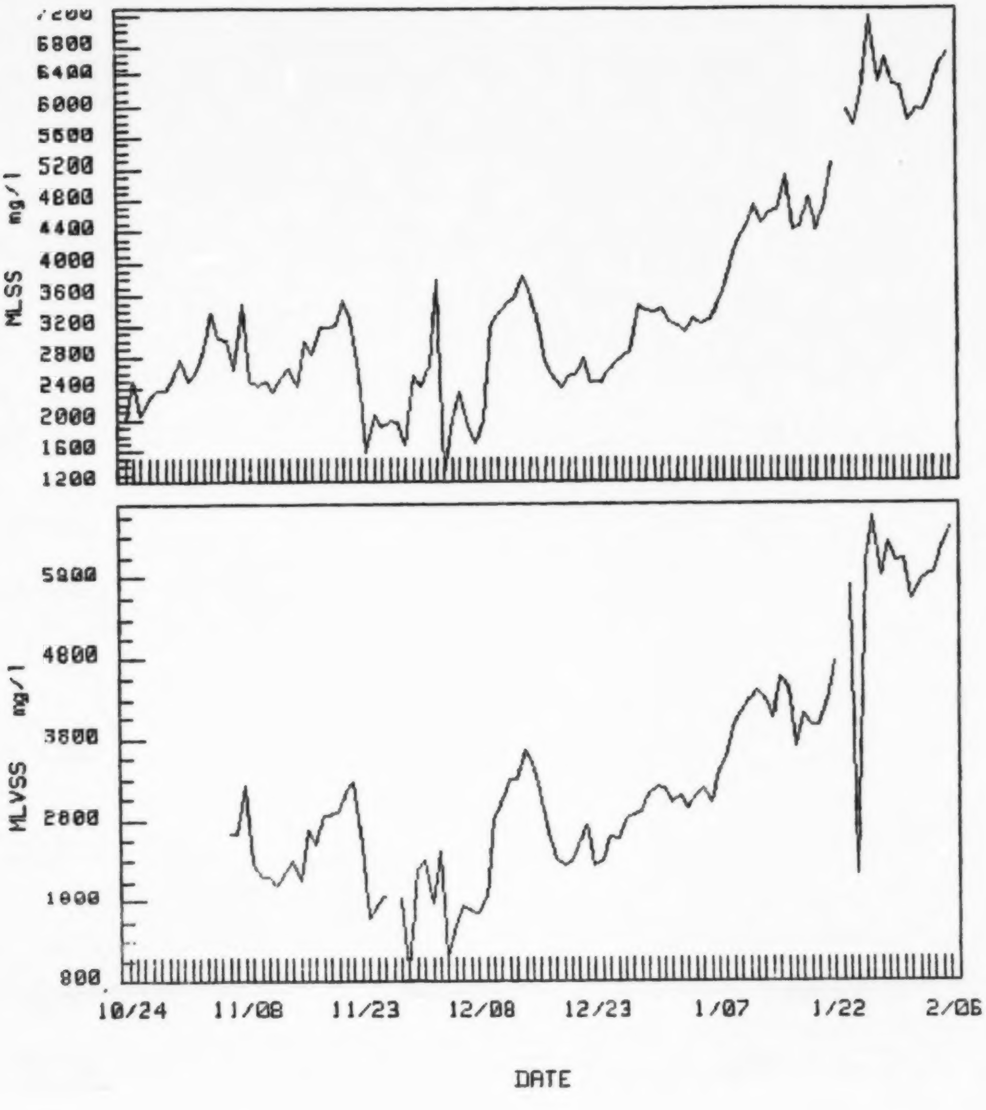
Many waste strength parameters, including phenol, NH_3, cyanide and BOD were found to be quite variable in the raw influent. This variability could be detrimental to treatment efficiency due to shock loading conditions and toxicity to organisms responsible for contaminant oxidation. In order to reduce the possibility of these occurrences two things were done. First, a low hydraulic loading was applied to the system in order to prevent shock to the biomass. The loading was maintained at approximately 3.785 liters/min (1 gpm) throughout the study. Second, raw influent was diluted prior to treatment in the same ratio used by the full-scale plant. The dilution ratio used was approximately 3 parts coke plant wastewater to 1 part service water. Use of this dilution and low hydraulic loading rates helped to maintain effective treatment through AS-1.

Nutrient Availability--

Efficient biological treatment is dependent on steady growth of the biomass and can be achieved by maintaining sufficient nutrients in the system. Nutrients of primary concern include phosphorus and nitrogen. During this study phosphorus in the form of phosphoric acid, was added to maintain desired phosphorus levels while nitrogen was present in sufficient quantities in the wastewater. These levels have been estimated in the literature in terms of ratios needed to remove certain contaminants. The ratio of BOD: Nitrogen:Phosphorus for effective BOD removal has been estimated at 100:5:1 (9). A phenol:phosphorus ratio of 70:1 has also been cited in the literature to facilitate phenol removal. A more complete discussion of the effects of nutrient deficiency and the food to microorganism (F/M) ratio will follow in the results section.

System Stabilization--

Due to the factors discussed above a certain amount of time was required before optimum treatment was attained. This stabilization can be seen in the plot of MLSS and MLVSS in Figure 19. Prior to about 12/14/79 the MLSS and MLVSS concentrations demonstrated great variability. After that date



(24 HR COMPOSITE SAMPLE DATA)

Figure 19. MLSS and MLVSS for AS-I aeration tank.

system upsets were minimized and the biomass was allowed to develop as shown by increasing MLSS and MLVSS concentration. As would be expected, treatment efficiency was not maximized until the system stabilized and a more complete discussion will follow.

Operational Results

The primary purpose of AS-I operation was to effectively remove phenol, organic carbon and thiocyanate. Each of these parameters and the factors that affected removal efficiency will be discussed below. A general table of influent and effluent mean and ranges for all parameters tested is shown in Table 24.

BOD--

Influent diluted feed BOD ranged from 1,290 mg/l to 2,550 mg/l and averaged 1,800 mg/l. Removal efficiency of BOD ranged from 60 to 99 percent and averaged 88 percent throughout the entire study. This removal efficiency, though, is not representative of the final results. During the final two-thirds of the study, after AS-I treatment had stabilized, BOD removal averaged 94 percent. This process stabilization period will be mentioned throughout this section and can again be visibly demonstrated in a plot of influent and effluent BOD and BOD removal as seen in Figures 20 and 21.

It is also apparent from the figures that effluent BOD and removal efficiency was dependent on influent BOD even after the system attained optimum removal. Complete influent and effluent BOD concentrations throughout the study can be found in Appendix B-3.4.

Phenol--

Excellent phenol removal was achieved through AS-I for the entire duration of the study. Influent (diluted feed) phenol concentrations ranged from 440 mg/l to 920 mg/l and averaged 657 mg/l. Treatment efficiency ranged from 90% - 100% with an average of 98.9%. Graphs of influent and effluent phenol concentrations and removal efficiency are shown in Figures 22 and 23.

On only three occasions did a significant disruption occur in the phenol removal efficiency. It should be noted that at these times effluent PO_4-P as measured by the lab and ortho-phosphate as P measured in the field were well below the recommended 70:1 phenol to phosphorus ratio. The corresponding decrease in the MLSS during these periods was also apparent, as expected. It should be noted that all of the upsets occurred prior to the aforementioned stabilization period. Influent and effluent phenol concentration and removal efficiencies can be found in Appendix B-3.5.

Thiocyanate--

Removal of thiocyanate during this study through AS-I was highly irregular. Influent thiocyanate ranged from 300 mg/l to 610 mg/l with an average of 450 mg/l while effluent concentrations ranged from 2 mg/l to 525 mg/l. This is illustrated in Figure 24. There did not appear to be any significant correlation between thiocyanate removal and phosphorus concentration as previously reported in the literature (10), although just prior to the two

TABLE 24. COMPARISON OF INFLUENT VERSUS EFFLUENT VALUES ACROSS AS-I.

Parameter	AS-I[a] Influent		AS-I Effluent	
	Ave.	Range	Ave.	Range
pH, Units	7.55[b]	6.80-8.70	7.10[b]	5.70-8.40
SS, mg/l	69	20-205	127	42-315
VSS, mg/l	39	19-86	76	39-158
$S^=$, mg/l	31	15-36	5.0	0.8-23
TDS, mg/l	7,540	4,770-10,000	7,830	5,500-9,000
Phenol, mg/l	655	440-920	8.3	<0.1-90
O&G, mg/l	17	2-105	10	<1-57
TKN, mg/l	784	330-2,010	772	360-1,600
NH_3 as N, mg/l	772	220-2,090	720	330-1,750
CN_T, mg/l	9.9	0.8-32	9.2	0.2-25
CN_A, mg/l	1.1	<0.05-12	0.96	<0.05-9.2
TOC, mg/l	1,110	180-1,660	242	35-740
TP as P, mg/l	9.71	<0.02-20.1	7.3	1.0-16
BOD, mg/l	1,800	1,290-2,550	215	13-905
Color, APHA Units	525	150-2,000	1,000	400-1,600
SCN, mg/l	450	300-610	323	2-525
NO_2/NO_3, mg/l	2	<1-5	2	1-2
SO_4, mg/l	1,600	-	1,980	-
Cl, mg/l	4,500	-	3,700	-
Na, mg/l	2,050	1,700-2,400	1,800	-
K, mg/l	5.4	5.3-5.4	5.0	-
Ca, mg/l	2.2	1.9-2.5	1.5	-
Mg, mg/l	4.3	3.4-5.2	5.8	-
Mn, mg/l	0.3	<0.1-0.4	0.4	-
Fe, mg/l	5.2	2.9-7.5	2.2	-
Alk., as $CaCO_3$, mg/l	325	210-535	665	104-1,000

a. Diluted 3 parts coke plant wastewater with 1 part service water.
b. Typical value.

NOTE: Service water quality is shown in Appendix A-2.

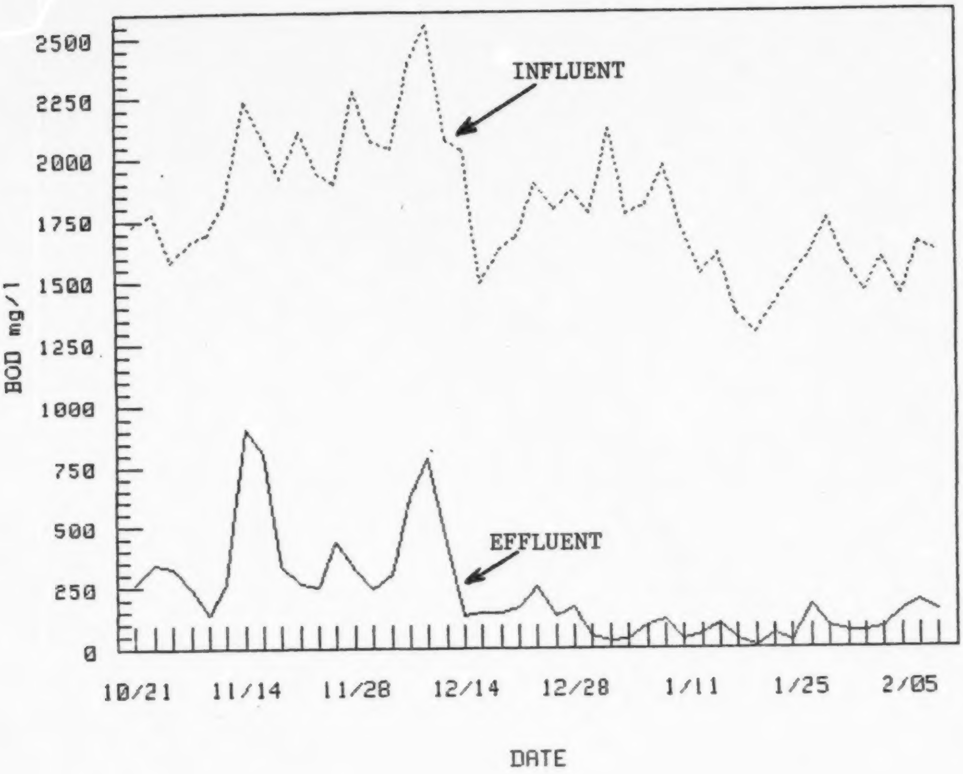


Figure 20. Influent versus effluent BOD for bioxidation unit AS-1.

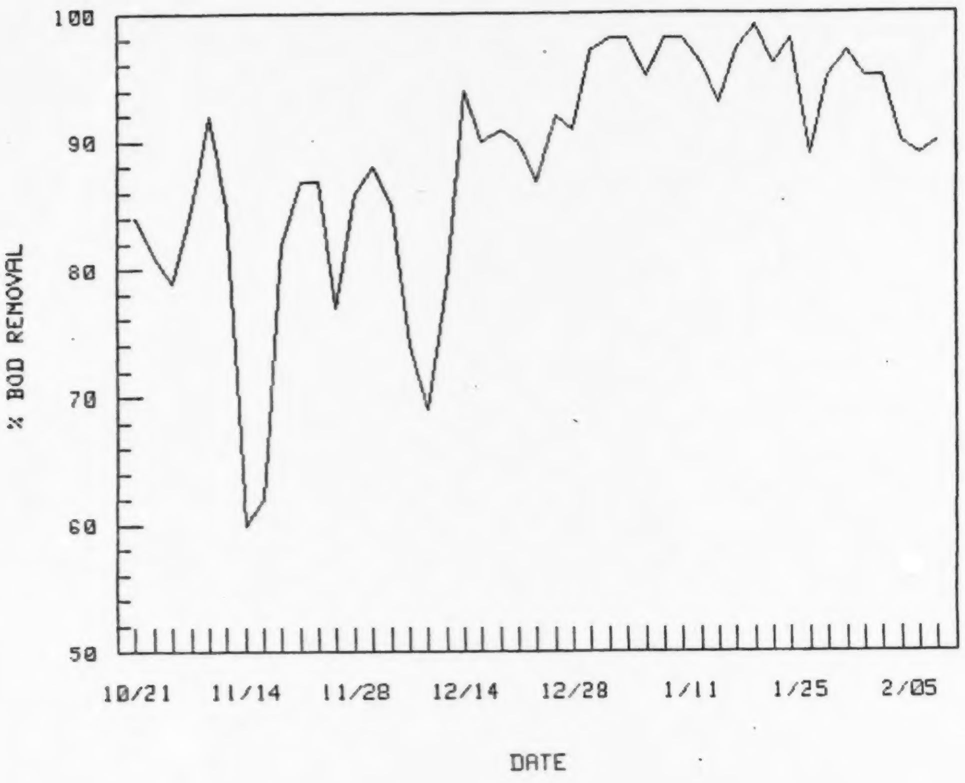


Figure 21. Percent BOD removal achieved across bio-oxidation unit AS-I.

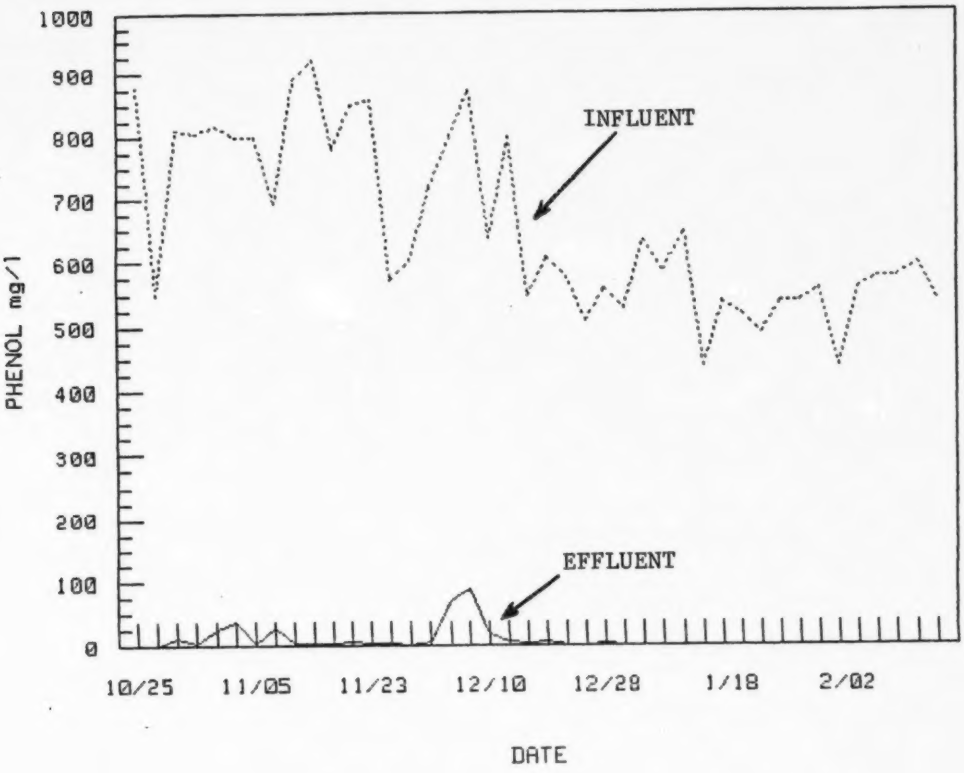


Figure 22. Influent versus effluent phenol for bioxidation unit AS-I.

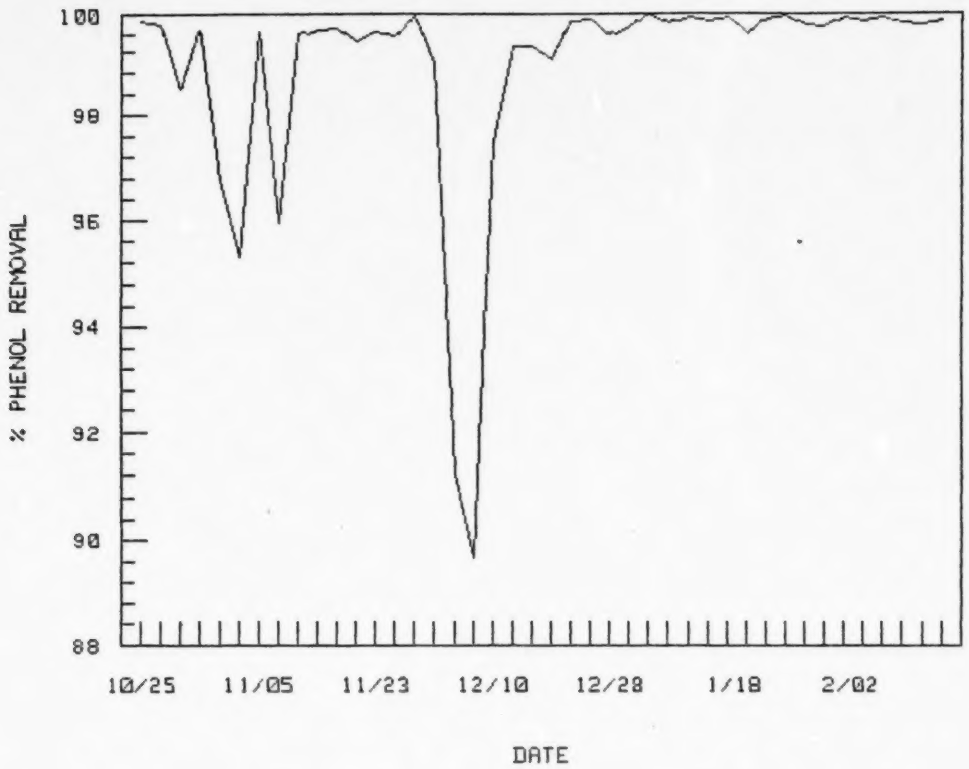


Figure 23. Phenol removal achieved across bio-oxidation unit AS-1.

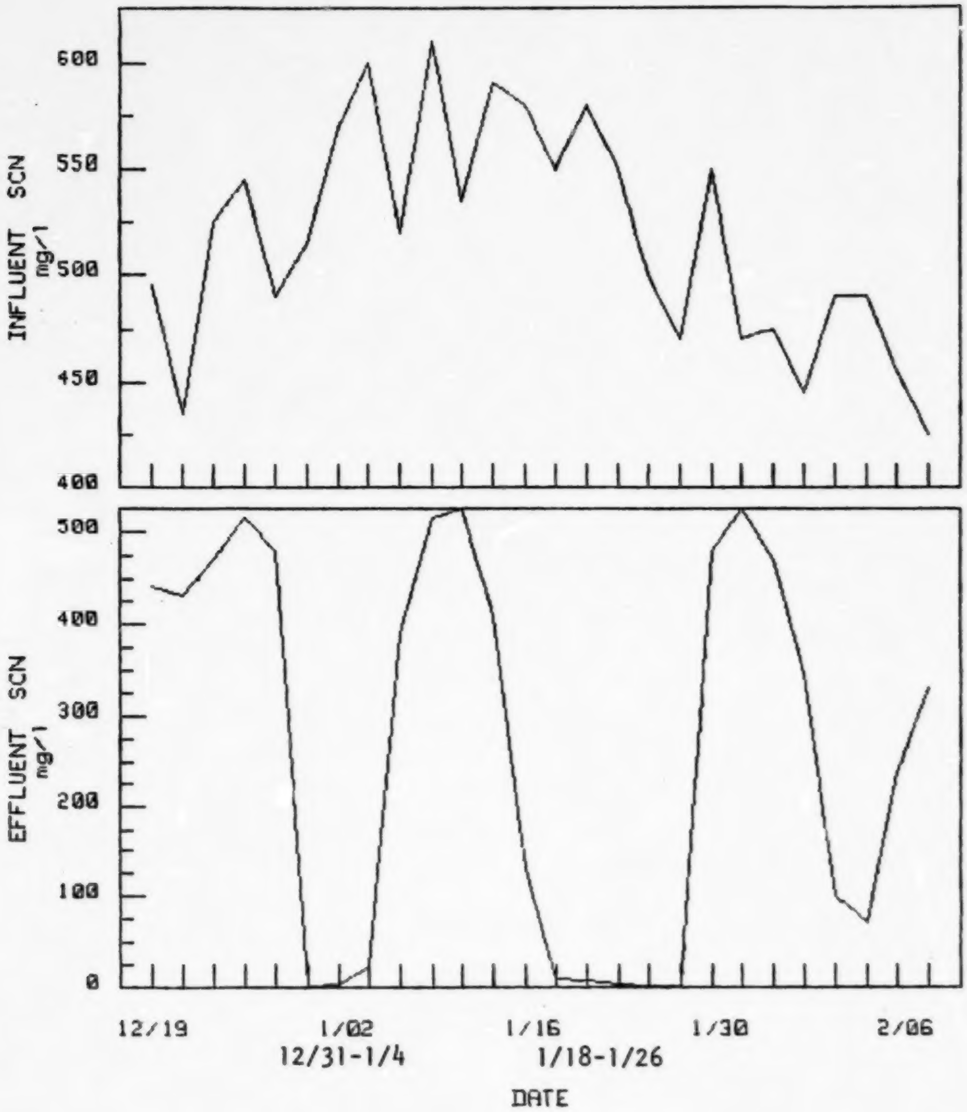


Figure 24. Influent and effluent thiocyanate concentrations - AS-1.

periods of good thiocyanate removal (12/13-1/4 and 1/18-1/26) measurements of both lab and field phosphorus reached higher than normal levels. This is shown in Figure 25. Complete thiocyanate and phosphorus measurements can be found in Appendix B-3.6.

Other Treatment Results--
Although little or no removal of a number of other contaminants was expected, it is interesting to note what did happen. Referring back to Table 24 at the beginning of this section it can be seen that no solids removal occurred across AS-1. A high degree of variability was also evident in the effluent suspended solids data as illustrated in Figure 26.

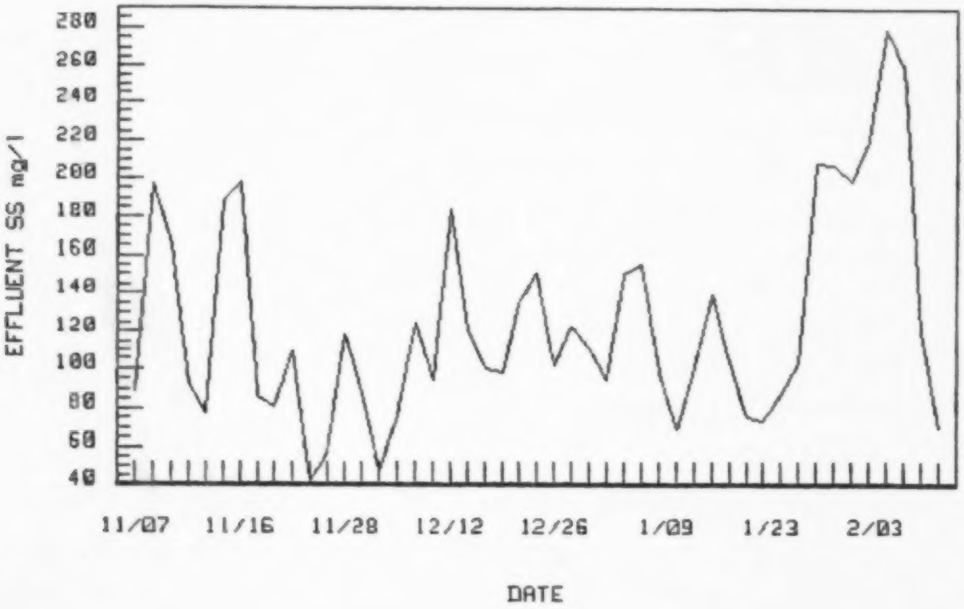


Figure 26. Effluent suspended solids concentration from AS-1 clarifier.

Two different determinations of nitrogen, namely ammonia nitrogen (NH_3 as N) and Total Kjeldahl Nitrogen (TKN) were removed in negligible amounts. Removal efficiencies for NH_3 and TKN were 6.7 percent and 1.5 percent, respectively. Total cyanide (CN_T) and cyanide amenable to chlorination also demonstrated small removal efficiencies of 7.1 percent and 12.7 percent, respectively. As would be expected Total Organic Carbon (TOC) was removed somewhat more effectively. Influent values and effluent concentrations averaged 1,110 mg/l and 242 mg/l, respectively, yielding a removal efficiency of 78 percent.

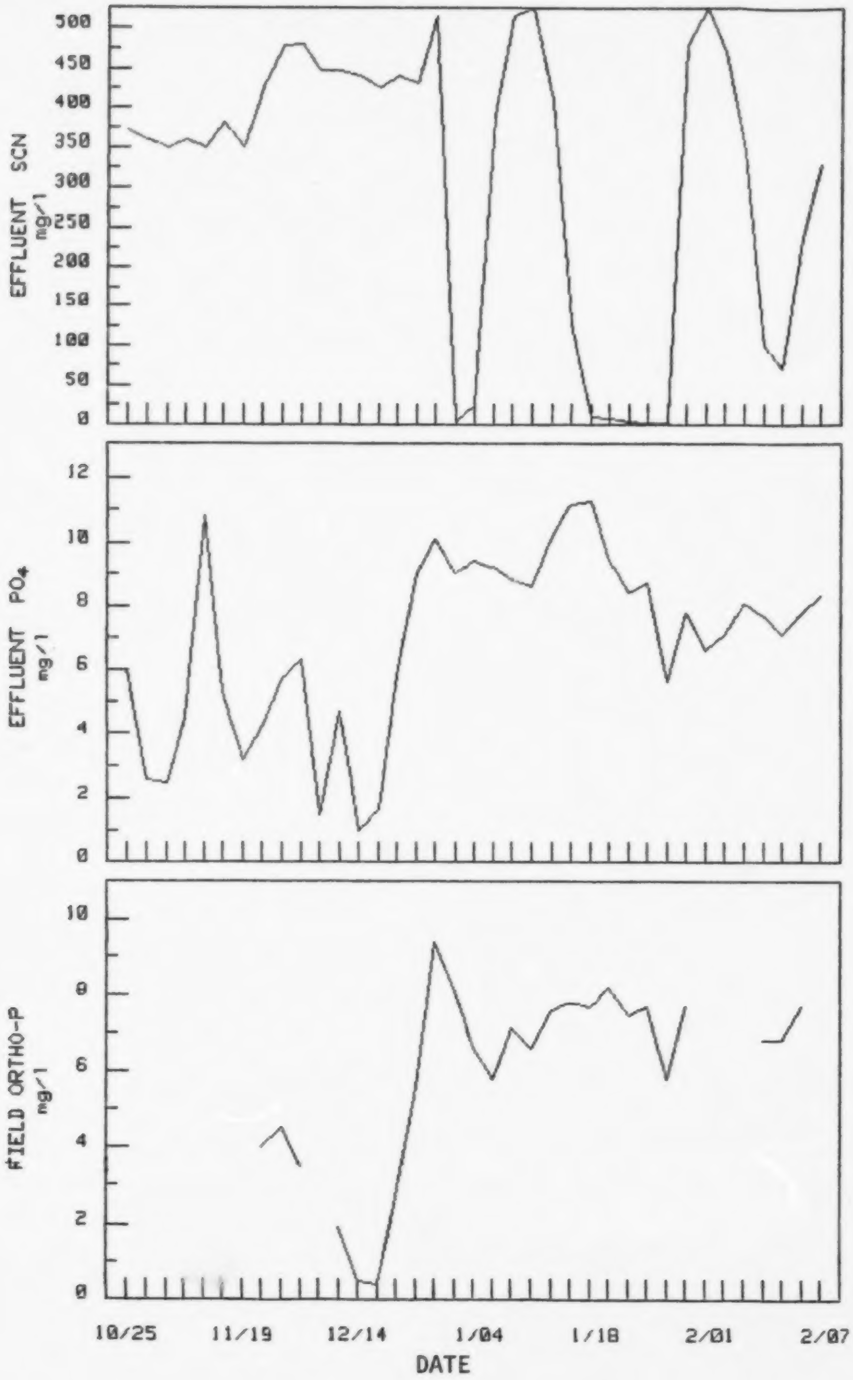


Figure 25. Effluent thiocyanate, laboratory phosphate and field phosphorus measurements - AS-1.

The biological solids encountered during this study demonstrated poor settling characteristics. In order to improve settling it was necessary to add a high charge cationic polyelectrolyte. A polymer dosage of 5 mg/l was used to maintain the sludge volume index (SVI) at approximately 100 or less. For the mixed liquor suspended solids (MLSS) concentrations encountered, a SVI of 100 represents a rough transition point between good and poor settling sludge. The settling rate was also measured regularly to help determine settling characteristics and polymer dosage. Ranges, average values and standard deviations found in the SVI and settling rate are listed below in Table 25. It did not appear that the settling rate or SVI was dependent on either MLSS or MLVSS.

TABLE 25. SVI AND SETTLING RATE DATA

Parameter	Range	Mean	Std. dev.
SVI	18-155	87	32
Settling rate	0.01-0.22 ft/min	0.09	0.05

Note: ft/min x 0.3048 = m/min

Daily information on MLSS, MLVSS, SVI, settling rate and return sludge solids concentration is shown in Appendix B-3.7.

Food to Microorganism Ratio (F/M)--
The food to microorganism ratio (F/M) maintained in an aeration tank defines the effectiveness of an activated sludge system. During the first part of the study the F/M ratio was highly variable and greater than desired. Once the stabilization period ended the F/M ratio decreased and was more consistent. This is illustrated in Figure 27 with F/M plotted against time.

The variability was due to a number of factors. Early in the study higher than normal levels of phenol and BOD provided excess food for the organisms. This food could not be entirely utilized, though, since lower than required levels of nutrients and dissolved oxygen prevented development of the biomass, as seen by referring to Figures 18 and 25. Here it should be noted that MLSS and MLVSS did not start increasing at a steady rate until the F/M was controlled. When the F/M increased to above 0.4 day^{-1} the organisms were in dispersed growth resulting in poor settling characteristics and poor BOD removal. High F/M ratios also have a negative effect on phenol and thiocyanate removal.

A number of occurrences combined to reduce the F/M and promote endogenous metabolic activity. First, BOD and phenol concentrations in the influent dropped rapidly (~12/17/79) due to coke plant wastewater treatment changes which reduced food to the biomass. When combined with increased nutrient

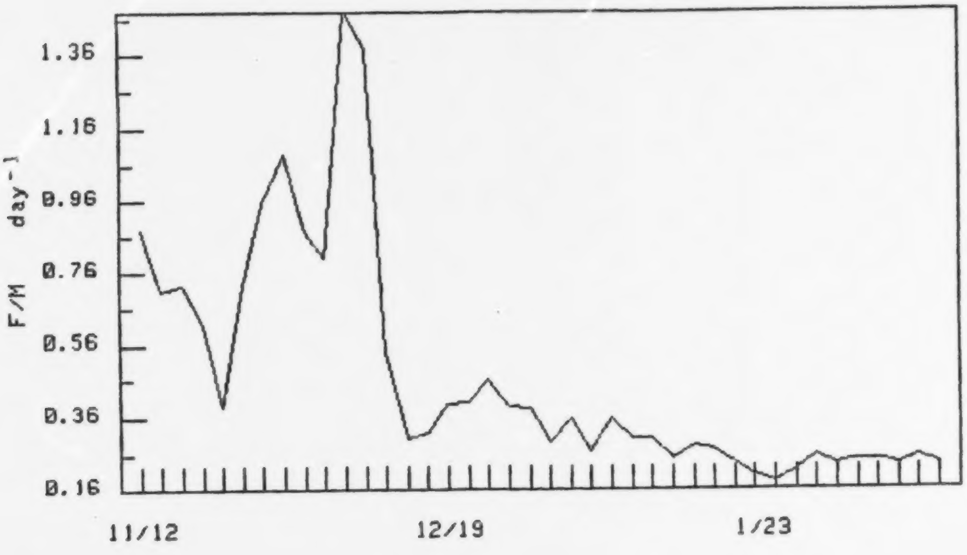


Figure 27. F/M versus time - AS-I.

levels the F/M was brought down to a more acceptable level (12/10/79 - 12/17/80). This aided microorganism growth until a minor upset was caused by a sudden drop in pH and temperature (12/17/79) which resulted in a drop in MLSS and MLVSS. This drop was also facilitated by some wasting of return sludge from AS-I. This upset, though, did not significantly affect continued stable operation. At this point (12/17/79 - 2/7/80) the biomass had become established and the MLSS and MLVSS continued to rise. Removal of BOD and phenol increased and was maintained through the study.

In order to more closely examine the effect of changing F/M ratios, four operating periods were selected during the study in which a relatively steady F/M ratio and an established biomass were maintained. The four selected periods had average F/M ratios of 0.24 day^{-1}, 0.413 day^{-1}, 0.735 day^{-1}, and 0.917 day^{-1}, respectively. Shown in Table 26 are average data taken during these periods. Complete data for these periods are found in Appendices B-3.8 and B-3.9.

This table demonstrates that a number of things are affected by the F/M ratio. As F/M increases, BOD removal generally decreases. The same could be said for phenol removal although these data do not clearly show it. Although the SVI does not appear to be directly related to F/M, the settling rate definitely is.

TABLE 26. FOUR SELECTED PERIODS OF CONSISTENT OPERATION FOR AS-I

Period	Ave. F/M, day^{-1}	Ave. BOD removal, %	Ave. Phenol removal, %	Ave. SVI	Ave. Settling rate, ft/min
11/23-11/28	0.917	83.4	99.7	64	0.067
11/12-11/19	0.735	71.3	99.6	120	0.062
12/19-12/26	0.413	90.0	99.6	48	0.036
1/30-2/7	0.24	93.0	99.9	106	0.033

Note: ft/min x 0.3048 = m/min

Special Testing--

During the final phase of the study when the system was operating at steady-state, testing was undertaken to determine oxygen uptake rates at various F/M ratios and the effectiveness of gravity thickening. Mixed liquor was combined in various amounts with raw waste in order to vary the F/M ratio during the oxygen uptake tests. This demonstrated that at increasing F/M ratios, oxygen uptake rates also increased. These data can be seen in Table 27.

TABLE 27. OXYGEN DEMAND AT VARIOUS F/M RATIOS - AS-I

Date	Time	Location	MLVSS (mg/l)	BOD (mg/l)	O_2 Demand (mg/l/hr)	F/M Ratio (kg BOD/day/ kg MLVSS)
2/4/80	1400	AS-I	4724	875	191	0.17
2/4/80	1430	AS-I	2386	875	230	0.34
2/4/80	1500	AS-I	1612	875	269	0.50
2/4/80	1530	AS-I	1223	875	309	0.66
2/6/80	1400	AS-I	5309	874	185	0.15
2/6/80	1430	AS-I	2671	874	193	0.30
2/6/80	1500	AS-I	1798	874	208	0.44
2/6/80	1530	AS-I	1359	874	217	0.59

Oxygen uptake tests were also conducted on each of the three aeration cells of AS-I to obtain background information on the oxygen demand of each cell. These data, which are presented in Appendix B-3.10 indicated that uptake in cell 1 was initially high (306 mg/l/hr on 2/3/80) and decreased as the waste passed through the system (103 mg/l/hr for cell 2 and 42 mg/l/hr for cell 3 on 2/3/80). The information shown in Appendix B-3.10 was used in conjunction with that previously presented in Table 27 to design the full-scale aeration systems.

Gravity thickening tests were also performed during the intensive test period with the system operating at steady state conditions. The tests were conducted on the return sludge from the first stage activated sludge clarifier on 2/3/80 and 2/7/80. Various dilutions were made on the sludge and settling rates were recorded. The settling data used in preparing the flux concentration curves are listed below in Table 28. The graphed flux concentration curves are shown in Appendices B-3.11 and B-3.12.

TABLE 28. SETTLING DATA FOR FLUX CONCENTRATION CURVE - AS-I

Source	Date	Sludge conc. (mg/l SS)	Settling rate (m/hr)
Return Sludge-1	2/3/80	6,000	0.088
Return Sludge-1	2/3/80	8,000	0.043
Return Sludge-1	2/3/80	10,000	0.024
Return Sludge-1	2/3/80	12,000	0.0152
Return Sludge-1	2/3/80	14,000	0.0103
Return Sludge-1	2/3/80	16,000	0.0073
Return Sludge-1	2/3/80	18,000	0.0059
Return Sludge-1	2/3/80	20,000	0.0047
Return Sludge-1	2/7/80	6,000	0.057
Return Sludge-1	2/7/80	7,000	0.0325
Return Sludge-1	2/7/80	8,000	0.023
Return Sludge-1	2/7/80	10,000	0.0133
Return Sludge-1	2/7/80	12,000	0.0082
Return Sludge-1	2/7/80	16,000	0.0041
Return Sludge-1	2/7/80	21,000	0.0022

From Appendices B-3.11 and B-3.12, it was determined that a full scale gravity thickening system could routinely achieve an underflow suspended solids concentration of 20,000 mg/l. This would result in a mass loading of 11.0 kg/m^2/day (2.25 lb/ft^2/day). Higher loadings would result in reduced thickener underflow concentrations, while a decrease in mass loading will increase the thickener underflow. This is shown in the following table.

MASS LOADINGS AT VARIOUS UNDERFLOW CONCENTRATIONS

Source	Date	Thickener underflow conc. (mg/l SS)	Maximum mass loading (kg/day/sq m)
Return Sludge-1	2/3/80	16,000	15.5
Return Sludge-1	2/3/80	18,000	13.1
Return Sludge-1	2/3/80	20,000	11.0
Return Sludge-1	2/3/80	22,000	9.5
Return Sludge-1	2/3/80	24,000	8.2
Return Sludge-1	2/7/80	16,000	8.5
Return Sludge-1	2/7/80	18,000	7.0
Return Sludge-1	2/7/80	20,000	5.9
Return Sludge-1	2/7/80	22,000	5.1
Return Sludge-1	2/7/80	24,000	4.5

Full Scale System Design, Costs, and Space Requirements

Introduction--

The design of a full scale wastewater treatment plant depends on a select number of process parameters. The costs of full-scale systems depend on aeration tank volume, air requirements, clarifier area, sludge production rate, etc. These system requirements are directly related to process parameters such as influent BOD and ammonia concentrations, mixed liquor suspended solids and mixed liquor volatile suspended solids (MLSS and MLVSS), food to mass (F/M) ratios, oxygen uptake rates, and settling rate.

Secondary process parameters affect the selection of the operating range for the primary process parameters. Some secondary process parameters are: phenol, thiocyanate, phosphorus concentrations, BOD/TKN ratio, and sludge age. Limits on the ratios of secondary parameters to the primary parameters may be found in the literature. The designs used for cost purposes are based on the pilot scale data and the limits found in the literature.

To generate design data, the pilot system was operated during extended periods under stable operating conditions. The operating conditions were changed for each extended period to evaluate performance over a range of conditions. By operating the pilot system for long periods under stable conditions, the average operating data became reliable. The data established by operating over a wide range of parameters can be used to predict full-scale performance at operating conditions within the range of the test conditions.

Some full-scale design data were obtained from bench scale tests. For example, settling tests were performed to determine settling rate and sludge thickening characteristics. A respirometer was used to determine typical oxygen uptake rates. Since the bench scale tests were performed with samples taken from the pilot system, the data are representative for the operating conditions existing at that time.

To select test data for design purposes, the MLSS data (Figure 19) and the field log books were examined. Stable periods of operation, as shown graphically on Figure 19, were carefully compared to the log book entries for those periods. Data were selected for design purposes only from those periods of operation when the following conditions were met:

1. The flow was constant.
2. Sludge was not wasted.
3. There were no upsets, such as accidental sludge wasting, foam-overs, poor settling quality, etc.
4. There were no unusual circumstances.

The data were examined and selected only if the data possessed the following characteristics:

1. The F/M ratio was within the selected range.
2. The feed water BOD or ammonia concentrations were within the range for that period.
3. The MLSS concentration was increasing at a stable rate.

After careful examination, four distinct periods of operation were selected during which the data were consistent with the above requirements. Other performance indicators, such as SVI, settling rate, and effluent quality also influenced the choice of operating periods.

Two wastewater treatment systems were designed. Design No. 1 is based on pilot plant data and the condition of wastewater at the test site. Design No. 2 is also based on pilot plant operating data, but assumes the wastewater feed is one that is expected to result from a properly operated pretreatment system (i.e. fixed and free ammonia stills). Characteristics of each type of wastewater are shown in Table 29. The table shows the BOD loading will be 50 percent less for a coke plant which discharges wastewater from a properly operating ammonia still (Design No. 2). The full-scale system for Design No. 1 requires dilution water, therefore, capital and operating costs will be greater to process the increased volume of water.

TABLE 29. STAGE I FEEDWATER CHARACTERISTICS FOR CARBONACEOUS REMOVAL SYSTEM DESIGNS - AS-I

Parameter	Design No. 1[a] site conditions	Design No. 2 proper ammonia still operation
Influent ammonia, mg/l	700	200
Influent BOD, mg/l	2,000	1,000
Design effluent BOD, mg/l	200	100
Design BOD removal, percent	90	90
Design effluent SS, mg/l	54	54
Design settling rate, ft/min	0.036	0.036

Note: ft/min x 0.3048 = m/min

a. Assumes pre-dilution of wastewater (3 parts coke plant wastewater to 1 part service water).

TABLE 30. AVERAGE OPERATING CONDITIONS FOR STAGE I DURING FOUR SELECTED PERIODS

Parameter	Test Period			
	11/12-11/19	11/23-11/28	12/19-12/26	1/30-2/7
Influent BOD, mg/l[a]	2,020	2,039	1,750	1,584
Effluent BOD, mg/l	580	337	175	111
MLVSS, mg/l	2,432	1,900	2,424	5,981
Detention time, hours	27	28	42.5	26.2
Flow, liters/min	3.7	3.6	2.3	3.77
F/M ratio, day^{-1} kg BOD/day/kg MLVSS	0.735	0.917	0.413	0.24
SVI	120	64	48	106
Settling rate, ft/min	0.062	0.067	0.036	0.033
Effluent SS, mg/l	74	53	34	89[b]

Note: ft/min x 0.3048 = m/min

[a] Feedwater consists of prediluted wastewater (3 parts coke plant wastewater to 1 part service water).

[b] No polymer was added during this period.

For both designs, a pre-mix system would be required, which would include a completely mixed tank and provision for chemical addition (phosphoric acid, sulfuric acid and caustic soda). As mentioned previously, dilution water for Design No. 1 would be required and added directly to the mixing tank in the ratio of 3 parts coke plant wastewater to 1 part dilution water. There would be no dilution water requirement for Design No. 2.

Choice of Steady State Periods for Design and Characterization of Performance--

The first stage of the pilot system was operated under four distinct sets of conditions. Variation of MLVSS was controlled to maintain the desired F/M ratios. Data from these four periods are summarized in Appendices B-3.8 and B-3.13. The MLVSS variation was plotted, and was shown previously in Figure 19. The average Stage I operating conditions for each period are shown in Table 30.

Stage I Design Criteria--

Selection of the MLVSS concentration is an important step in the design of a biological wastewater treatment system. The MLVSS concentration directly affects other process parameters such as F/M ratio, hydraulic retention time (HRT), sludge production, and sludge age (solids retention time, SRT). The range of acceptable MLVSS has been reported to be between 2,000-12,000 mg/l for coke plant wastes (11). Some investigators (12, 13, 14) have used MLVSS concentrations in the range of 2,500-5,700 mg/l successfully with similar water.

The food to mass ratio (F/M) is also an important design parameter. Values of F/M have been reported up to 0.4 kg of BOD/day/kg of MLVSS for biological systems with adequate nutrients (13). For systems combining carbonaceous removal and nitrification in one aeration basin, the limiting F/M is 0.3 kg BOD/day/kg MLVSS (13). Separate systems may operate at F/M ratios up to 0.5 day^{-1}, since the nitrifiers in the second stage will not be affected by the more active organic removing organisms present in the first stage (12, 13).

Hydraulic retention time, HRT, is defined as:

$$\text{HRT} = \frac{V}{Q} \qquad \text{where}$$

V is the aeration tank volume, Q is the waste flow. The literature reports that HRT ranges up to 35 hours with 24 hours being common (11) for coke plant wastewater treatment systems. The HRT of a system varies inversely with MLVSS concentration and the F/M ratio (15).

Sludge production is also affected by MLVSS concentration. A high MLVSS concentration will increase endogenous respiration, thereby reducing sludge solids production (13). However, excessive oxygen will be required that is not used for the removal of organic compounds.

Sludge age of up to 80 days has been reported for bench test investigations, however, 30 days is recommended by one investigator (14). Other investigators recommend sludge age of 6-12 days (3, 16). The definition of sludge age as used in this discussion is the total sludge solids in aeration divided by total solids lost per day.

From the above discussion, it can be seen that the parameters used in the design of the first stage aeration system are interrelated. For design purposes, a MLVSS concentration of 3,500 mg/l was ultimately selected. The pilot plant data showed that operation at a F/M ratio of 0.413 kg BOD/day/kg MLVSS would result in 90 percent BOD removal. Effluent suspended solids and SVI were low at that F/M ratio, indicating satisfactory settling characteristics. Therefore, the F/M ratio of 0.4 kg BOD/day/kg MLVSS was selected for design purposes. The other design parameters also would be within the range of values recommended in the literature. All design parameters are summarized in Table 31.

TABLE 31. STAGE I DESIGN PARAMETERS

Parameter	Recommended range	Design No. 1	Design No. 2
MLVSS, mg/l	2,000-12,000	3,500	3,500
F/M, kg BOD/day/kg MLVSS	<0.5	0.4	0.4
HRT, hours	24-35	34	29
SRT, days	6-12	7.4	8.1
Sludge production, kg per day (for 3,785 m^3/day plant)	-	2,812	2,139
Recycle rate, percent	-	42	44.4

Secondary design factors considered were: thiocyanate ratio; phenol ratio; phosphorus ratio; and BOD/TKN ratio. The literature states that the thiocyanate should not exceed 0.14 times the MLVSS concentration (12). Phenol concentration should not exceed 0.25 times the MLVSS concentration (17). The ratio of phosphorus to phenol concentrations should be at least 1:70 (12). The ratio of Stage I effluent BOD to TKN should be in the range of 0.5-1.0 (18) for good nitrification, or 2-4 for good floc settlement in Stage II. These ratio requirements were all met with the selected design parameters and predicted performance data, as shown below in Table 32.

TABLE 32. SECONDARY DESIGN FACTORS

Parameter	Recommended maximum	Design No. 1	Design No. 2
Thiocyanate to MLVSS ratio	0.14	0.109	0.109
Phenol to MLVSS ratio	0.25	0.184	0.184
Phosphorus to phenol ratio	1:70	>1:70[a]	>1:70[a]
BOD/TKN	2-4	2.8	5

a. Necessary phosphorus would be added, on the basis of pilot plant data, in the form of phosphoric acid.

Other important design data were extrapolated from the pilot plant data. These data, which were used in the design calculations, are summarized in Table 33.

Equipment Description, Costs, and Space Requirements (Mixing System)--
The coke plant investigated had an equalization system consisting of three tanks in series. The equalized flow was then fed to the biological treatment system. The forthcoming designs all assume that an equalized feed flow is available. A full-scale mixing system, as shown previously in Figure 12, should be provided for each design. Dilution water is assumed to be required at the rate of one part per every three parts of coke plant effluent, in Design No. 1. This is the dilution ratio currently used in the full-scale treatment plant which ran in parallel to the pilot system.

In Design No. 2, no dilution water has been assumed in the design, and it is unknown if it would be needed in full-scale operations.

Dilution water would also be required to reduce the Stage 1 effluent ammonia to approximately 200 mg/l, in Design No. 1. No dilution water is required on the Stage 1 effluent of Design No. 2, since the coke plant effluent would contain about 200 mg/l of ammonia if the ammonia stills were operating properly. Further discussion of this is presented in the following section.

The mixing system has been designed for effluent flows of 378 m^3/day (1.0 mgd), 1,892 m^3/day (0.5 mgd) and 3,785 m^3/day (1.0 mgd). Detention time has been set at 1 hour. For Design No. 1, the influent wastewater flow would be approximately 75 percent of the effluent flow, since dilution water makes up the remaining 25 percent. For Design No. 2, the influent flow

TABLE 33. PILOT PLANT DESIGN DATA

Parameter	Pilot plant value
Settling rate, ft/min (mixed liquor)	0.036
Influent SS, mg/l	70
Effluent SS, mg/l	54
Oxygen required, mg/l	150
Return sludge maximum concentration, mg/l	10,000

Note: ft/min x 0.3048 = m/min

equals the effluent flow, since no dilution water has been assumed. It is assumed that a mixing system is required with both designs for chemical addition and pH adjustment. Although the mixing system could be made slightly smaller for Design No. 2, it is assumed that the cost difference would be negligible. The design basis for the mixing system is shown in Table 34.

Costs are summarized in Table 35. It should be emphasized that the mixing system costs must be added to the biological oxidation system costs to determine the complete costs of the system.

Equipment Description, (Stage I, Design No. 1)--
The carbonaceous removal system, Stage I, would consist of aeration tankage, clarification, sludge recycle system, blowers, and chemical feed systems. This equipment is shown by the diagram in Figure 28. Equipment sizes, quantity, and other descriptive information is shown on Table 36. Equipment costs are given in Table 37.

The aeration tanks would be built of cast-in-place concrete below ground. Each aeration tank would be fitted with fiberglass pipes with coarse bubble diffusers for air distribution. Air would be supplied to the aeration tanks by blowers. Table 36 lists the total air flow required for each size system. The number of blowers listed includes a spare blower which does not contribute to the total air supplied unless one of the other blowers fails.

Clarifiers would be built of cast-in-place concrete below ground. The clarifiers would be of the type having peripheral effluent weirs and rotating sludge collection manifolds.

TABLE 34. DESIGN BASIS FOR THE MIXING SYSTEM

Design flow, m^3/day (mgd)	378.5 (0.1)	1,892.5 (0.5)	3,785 (1.0)
Treatment tank size,[a] 1 hour detention time, m^3 (ft^3)	15.9 (560)	79 (2,800)	159 (5,600)
Chemical use, ℓ/day (gal./day)			
- acid	280 (74)	1,400 (372)	2,800 (740)
- caustic soda[b]	280 (74)	1,400 (372)	2,800 (740)
- phosphoric acid	21.9 (5.8)	109.8 (29)	219 (58)
Chemical storage For 30 day supply, ℓ (55 gal. bbl)			
- acid (98% H_2SO_4)	8,327 (40)	41,635 (200)	83,270 (400)
- caustic soda (25% NaOH)	208 (1)	1,041 (5)	2,082 (10)
- phosphoric acid (75% H_3PO_4)	625 (3)	3,331 (16)	6,453 (31)
Connected power, kw (hp)	3.92 (5.25)	13.62 (18.25)	26.86 (36)

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a. Assuming 25% of the mixture is dilution water.
b. Caustic soda use is negligible, however it is required occasionally to increase pH.

TABLE 35. COST SUMMARY, COKE PLANT BIO-OXIDATION MIXING SYSTEM

Design flow, m^3/day (mgd)	379 (0.1)	1,892 (0.5)	3,785 (1.0)
Capital Costs[a], $			
Mixing tank	$ 6,000	$ 13,000	$ 20,000
Mixer	3,000	7,000	10,000
Pumps (Feed & Chemical)	6,000	10,000	12,000
Acid storage tank	4,000	14,000	18,000
pH Controller	4,000	4,000	4,000
Electrical control panel	2,000	3,000	3,000
Delivered equipment total	$ 25,000	$ 51,000	$ 67,000
Equipment installation	10,000	20,000	27,000
Installed equipment total	$ 35,000	$ 71,000	$ 94,000
Process piping	10,000	21,000	28,000
Instrumentation	2,000	4,000	5,000
Capital equipment total	$ 47,000	$ 96,000	$127,000
Facilities & Fees	66,000	135,000	178,000
Capital cost total	$113,000	$231,000	$305,000
Operating Costs, ($/yr)			
Electrical	$ 1,000	$ 5,000	$ 10,000
Operator	4,000	4,000	4,000
Maintenance Labor & Materials	2,000	5,000	6,000
Chemicals	51,000	258,000	513,000
Yearly operating subtotal	$ 58,000	$272,000	$533,000
Capital Amortization			
20 yr life @ 7.125%	$ 11,000	$ 22,000	$ 29,000
Total Annualized Cost, $/yr[b]	$ 69,000	$294,000	$562,000
$/3,785 ℓ ($/1,000 gal.)	1.89	1.61	1.54
Land Requirements			
Equipment only, m^2 (ft^2)	38 (300)	84 (900)	139 (1,500)

a. See Appendices C and D for basis of estimate. All costs are effective as of December, 1980.

b. Annualized cost = operating cost + capital amortization.

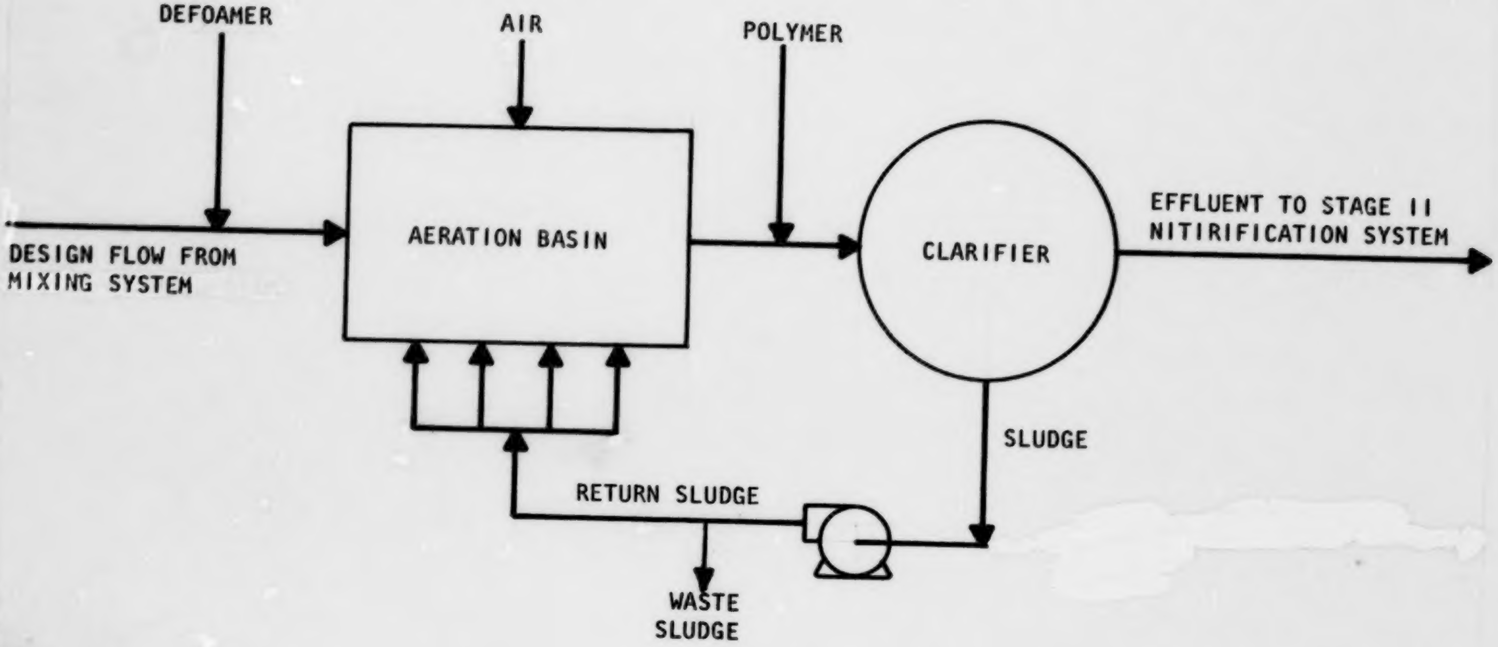

DEFOAMER
AIR
POLYMER
DESIGN FLOW FROM MIXING SYSTEM
AERATION BASIN
CLARIFIER
EFFLUENT TO STAGE II NITIRIFICATION SYSTEM
SLUDGE
RETURN SLUDGE
WASTE SLUDGE

Figure 28. Diagram of stage I completely mixed biological treatment system for coke plant wastewater (Design No. 1 and Design No. 2).

TABLE 36. EQUIPMENT DESCRIPTION - STAGE I - DESIGN NO. 1

Flow rate, m^3/day (mgd)	379 (0.1)	1,892 (0.5)	3,785 (1.0)
Total aeration volume, m^3 (ft^3)	540 (19,100)	2,700 (95,600)	5,400 (191,000)
Aeration tank number	2	2	3
Aeration tank size [wxlxh] m (ft)	7x9x5 (22x28x15)	9x30x5 (30x100x15)	9x43x5 (30x142x15)
Total air required, m^3/min (cfm)	106 (3,750)	531 (18,750)	1,060 (37,500)
Blower number	3	3	5
Defoamer[a], ℓ/day (gpd)	0.8 (0.2)	3.8 (1.0)	7.6 (2.0)
Polymer[b], ℓ/hr (gph)	87 (23)	416 (110)	870 (230)
Total clarifier area, m^2 (ft^2)	48 (515)	239 (2,577)	479 (5,154)
Clarifier number	1	2	2
Clarifier diameter, m (ft)	7.6 (25)	12.2 (40)	18.3 (60)

a. Based on 2 mg/l dosed.
b. Based on 5 mg/l dosed and 1 mg/ml stock.

TABLE 37. COST SUMMARY - STAGE 1 - DESIGN NO. 1

Design flow, m^3/day (mgd)	379 (0.1)	1,892 (0.5)	3,785 (1.0)
Capital Costs, \$[a]			
Wastewater pumps	\$ 7,000	\$ 8,000	\$ 9,000
Chemical pumps	4,000	5,000	6,000
Chemical mixers	1,000	1,000	2,000
Chemical tanks	1,000	2,000	3,000
Aeration tanks	30,000	102,000	187,000
Aeration piping	15,000	77,000	110,000
Aeration blowers	42,000	170,000	283,000
Clarifier	49,000	139,000	197,000
Electrical panel	11,000	30,000	38,000
Delivered equipment total	\$160,000	\$ 534,000	\$ 835,000
Equipment installation	64,000	214,000	334,000
Installed equipment total	\$224,000	\$ 748,000	\$1,169,000
Process piping	67,000	224,000	468,000
Instrumentation	11,000	37,000	58,000
Capital equipment total	\$302,000	\$1,009,000	\$1,695,000
Facilities & Fees	423,000	1,414,000	2,373,000
Capital cost total	\$725,000	\$2,423,000	\$4,068,000
Operating Costs, (\$/yr)			
Electrical	\$ 42,000	\$ 201,000	\$ 402,000
Operator	36,000	36,000	36,000
Maintenance labor & materials	15,000	50,000	85,000
Chemicals	2,000	11,000	22,000
Yearly operating subtotal	\$ 95,000	\$ 298,000	\$ 545,000
Capital Amortization			
20 yr life @ 7.125%	\$ 69,000	\$ 231,000	\$ 388,000
Total Annualized Cost, \$/yr[b]	\$164,000	\$ 529,000	\$ 933,000
\$/3,785 ℓ (\$/1,000 gal.)	4.49	2.90	2.56
Land Requirement			
m^2 (ft^2)	410 (4,400)	1,110 (12,000)	3,430 (36,900)

a. See Appendices C and D for basis of estimate. All prices are effective as of December, 1980.

b. Annualized cost = operating cost + capital amortization.

Sludge will be recycled from the clarifier back to the aeration tank by single pumps. A spare pump would be provided for emergency use in the event of a pump failure or for maintenance shutdowns. Waste sludge would be pumped to the sludge disposal system by a bleed valve on the return sludge line. There is no sludge disposal system included in the costs. Recycle system piping, valves, and controls are included in the costs.

The chemical supply system for this stage would consist of a defoamer pump and a polymer pump, with associated tankage and piping. Defoamer would be pumped at a rate equivalent to a dose of 2 mg/l. Polymer would be pumped at a rate equivalent to a dose of 5 mg/l. Spare pumps would be furnished for both defoamer and polymer. Defoamer would be pumped directly from the shipping barrels, while polymer would be pumped from a mix tank. The polymer mix tank would be provided with a mixer and enough capacity to require batch preparation only once per day.

The electrical control system would consist of a motor control center and a supervisory control console. Costs include these panels plus all conduits, wire, and installation labor.

A building would be required to contain the pumps, blowers, chemicals, controls, and a small laboratory space. The costs include building space for the equipment associated with this stage.

Equipment Costs--Using the design parameters developed in Tables 34 and 36, Stage I biological treatment system costs were computed for each of the design flows. Costs for system components were based on vendor quotes and from data in a recent edition of the Dodge Manual for Building Construction, Pricing, and Scheduling (7). Equipment costs include:

1. Concrete aeration tanks.
2. Aeration piping, diffusers, and blowers.
3. Concrete clarifier tanks.
4. Sludge collection system, piping, and pumps.
5. Chemical feed pumps, tanks, and piping.
6. Motor controls.

The Chilton factor method (8) was used to compute the costs of installation, process piping, instrumentation, facilities, and fees.

Operating Costs--
Components of the yearly operating cost include power, chemicals, operating labor, maintenance labor, and maintenance material. Items included in the analysis are discussed as follows:

a. Power costs were assumed to be 3.0¢ per KWH. It was assumed that all system motors, except the spare units, would be in operation 24 hours per day, 365 days per year.

b. Chemical consumption would be as listed in Table 36. Costs of chemical were based on bulk prices in effect on December 1, 1980. These prices were:

Polymer: $1.36/kg ($3.00/pound)
Defoamer: $1.45/liter ($5.50/gal)

c. Operating labor would be required to oversee the system continuously, regardless of size. It was assumed that a skilled operator earning $19,000 per year would be present during the first shift. Unskilled operators earning $16,000 per year would be present during the second and third shifts. The above wages are based on a 5 day work week. Since the system must be operated 7 days per week, additional labor costs were included for continuous operator presence. Assuming that a nitrification system will be used with this system, one half of the operator charges were allocated to this system.

d. Maintenance labor included routine inspection, repairs, and preventive maintenance.

e. Maintenance material allowances are included for replacement parts. This cost is a function of total capital costs.

Capital Amortization--An equipment service life of twenty years, amortized at 7.125 percent per year was assumed.

Items Not Included--Examples of items not included in this analysis are: fringe benefits, supervision, overhead, laboratory support, insurance, and administration expenses.

Cost comparisons are calculated on the basis of equivalent annual values which include both capital amortization and operating costs. Results shown in Table 37 are expressed as total yearly costs and costs per 3,785 liters (one thousand gallons) design flow.

Absolute Cost Consideration--
The cost estimates presented in Table 37 are internally consistent and suitable for comparison purposes within the scope of this report. It should be realized that the costs are limited to items listed previously and do not represent the absolute value of total installed cost. The intent is to compare the various alternatives with detail not available in published cost curves.

Land Requirements--
Also tabulated in Table 37 are the space requirements for the three designs. Space is provided for the major structures, such as aeration basins, clarifiers, and buildings, plus access space. The cost for facilities includes sidewalks and driveways which would occupy some of the access space around the structures.

Equipment Description, (Stage I, Design No. 2)--
The Stage I equipment for Design No. 2 would be of the same type as required for Design No. 1, however, sizes and quantities would decrease. All of the design calculations were recomputed on the basis of lower BOD concentration in the influent wastewater. Although the hydraulic flows are identical, tank sizes, power, and quantity of waste sludge could be reduced.

Equipment sizes, quantity, and other descriptive information are shown in Table 38. Equipment costs are given in Table 39. All of the previous discussion related to Stage I, Design No. 1, also applies to Design No. 2.

RESULTS OF THE ACTIVATED SLUDGE (NITROGEN REMOVAL) STUDY

Introduction

The discharge of ammonia, which is generated in the by-product cokemaking operation, is of primary concern because of its impact on lowering the receiving stream dissolved oxygen through the nitrification process. Through biological nitrification, ammonia is oxidized to nitrate with an intermediate formation of nitrite. The two principal microorganisms involved in nitrification are Nitrosomonas and Nitrobacter. These bacteria are autotrophic and derive energy from the oxidation of inorganic nitrogen compounds. An inorganic carbon source is also required for synthesis. Nitrifying bacteria have a slower growth rate than heterotrophic bacteria and therefore, require a longer solids retention time.

The oxidation of ammonia to nitrite by Nitrosomonas occurs according to the following reaction:

$$15CO_2 + 13NH_4^+ \longrightarrow 10NO_2^- + 3C_5H_7NO_2 + 23H^+ + 4H_2O$$

Nitrosomonas

The nitrite is then oxidized to nitrate by Nitrobacter as shown below:

$$5CO_2 + NH_4^+ \; 10NO_2^- + 2H_2O \longrightarrow 10NO_3^- + C_5H_7NO_2 + H^+$$

Nitrobacter

As seen in the above reactions, inorganic carbon (CO_2) is consumed while free acid (H^+) is produced. Thus, an alkalinity supplement must be provided to maintain a constant nitrification rate.

Separate stage nitrification was investigated during this study; that is, carbonaceous material was removed in the first stage activated sludge unit (AS-I). Thus, the organic loading on the nitrification activated sludge unit was relatively low compared to the influent ammonia load. This

TABLE 38. EQUIPMENT DESCRIPTION - STAGE I - DESIGN NO. 2

Flow rate, m^3/day (mgd)	379 (0.1)	1,892 (0.5)	3,785 (1.0)
Total aeration volume, m^3(ft^3)	460 (16,300)	2,300 (81,350)	4,600 (163,000)
Aeration tank number	2	2	3
Aeration tank size [wxlxh]m (ft)	7x7x5 (22x24x15)	9x27x5 (30x90x15)	9x37x5 (30x121x15)
Total air required, m^3/day (cfm)	90 (3,200)	450 (16,000)	900 (32,000)
Blower number	3	3	5
Defoamer[a], liters/day (gpd)	0.8 (0.2)	3.8 (1.0)	7.6 (2.0)
Polymer[b], liters/hour (gph)	87 (23)	416 (110)	870 (230)
Total clarifier area, m^2 (ft^2)	48 (515)	239 (2,577)	479 (5,154)
Clarifier number	1	2	2
Clarifier diameter, m (ft)	7.6 (25)	12.2 (40)	18.3 (60)

a. Based on 2 mg/l dosed.
b. Based on 5 mg/l dosed and 1 mg/ml stock.

TABLE 39. COST SUMMARY - STAGE 1 - DESIGN NO. 2

Design flow, m^3/day (mgd)	379 (0.1)	1,892 (0.5)	3,785 (1.0)
Capital Costs, \$[a]			
Wastewater pumps	\$ 7,000	\$ 8,000	\$ 9,000
Chemical pumps	4,000	5,000	6,000
Chemical mixers	1,000	1,000	2,000
Chemical tanks	1,000	2,000	3,000
Aeration tanks	26,000	89,000	163,000
Aeration piping	11,000	67,000	151,000
Aeration blowers	40,000	147,000	244,000
Clarifier	49,000	139,000	197,000
Electrical panel	11,000	26,000	32,000
Delivered equipment total	\$150,000	\$ 484,000	\$ 807,000
Equipment installation	60,000	194,000	323,000
Installed equipment total	\$210,000	\$ 678,000	\$1,130,000
Process piping	63,000	203,000	339,000
Instrumentation	10,000	34,000	56,000
Capital equipment total	\$283,000	\$ 915,000	\$1,525,000
Facilities & fees	396,000	1,281,000	2,135,000
Capital cost total	\$679,000	\$2,196,000	\$3,660,000
Operating Costs, (\$/yr)			
Electrical	\$ 35,000	\$ 172,000	\$ 344,000
Operator	36,000	36,000	36,000
Maintenance labor & materials	14,000	46,000	76,000
Chemicals	2,000	11,000	22,000
Yearly operating subtotal	\$ 87,000	\$ 265,000	\$ 478,000
Capital Amortization			
20 yr life @ 7.125%	\$ 65,000	\$ 209,000	\$ 349,000
Total Annualized Cost, \$/yr[b]	\$152,000	\$ 474,000	\$ 827,000
\$/3,785 ℓ (\$/1,000 gal.)	4.16	2.60	2.27
Land Requirement			
m^2 (ft^2)	700 (2,300)	5,420 (17,800)	10,100 (33,100)

a. See Appendices C and D for basis of estimate. All prices are effective as of December, 1980.

b. Annualized costs = operating cost + capital amortization.

resulted in a higher population of nitrifiers and corresponding higher rates of nitrification. During the study, the first stage activated sludge unit's clarified effluent was used as a source of feed as shown in Figure 29.

The investigation was performed from October, 1979 to February, 1980. Presented below is a detailed discussion of the pilot study findings pertaining to the second stage activated sludge unit's performance.

General Results

Start-up of Nitrification System--

During start-up, the nitrification system tanks were filled with water as part of the on-site wet test of the equipment. On the second day of operation, the clarifier from the first stage activated sludge unit started to overflow at a rate of 3.785 ℓ/min (1 gpm). At this time, a slurry of 29.5 kg (65 lb) of powdered activated carbon (PAC) was added to the aeration tank. This was equivalent to a dosage of 5,000 mg/l. The blowers were turned on to keep the carbon in suspension. The powdered carbon was added to help in settling the biomass and to reduce the effects of any toxics passing through the system. The sludge recycle pump was turned on causing flow from the clarifier to the aeration tank. The pump was set so that little or no sludge blanket formed in the clarifier. Polymer was added to the clarifier to improve settling and reduce the loss of biomass over the weir. Bench screening tests were performed to optimize polymer type and dosage. A high density cationic polymer was found to be optimum. About 10 grams (0.35 oz) of a commercially available mutant bacteria strain were added to the aeration tank to promote the formation of nitrifying organisms. Sodium bicarbonate was also added in amounts sufficient to keep the pH in the 7.8-8.2 range and maintain a clarifier effluent alkalinity of about 100 mg/l. Daily field analyses were performed, including mixed liquor dissolved oxygen, temperature, mixed liquor suspended solids (MLSS), mixed liquor volatile suspended solids (MLVSS), effluent pH, alkalinity and ammonia. The settled sludge from the secondary clarifier continued to be recirculated without wasting for a period of about 6 weeks without any significant growth of nitrifying organisms. During this period, influent ammonia was quite variable, averaging 1,050 mg/l with a range from 620 mg/l to 1,750 mg/l. This is shown graphically in Figure 30. A summary of the parameter concentrations in the second stage activated sludge unit during the period October 22 - November 29, 1979 is presented in Table 40. It is evident from this table that no ammonia was removed during this period. A comprehensive listing of all analytical research data taken during the entire study can be found in Appendix B-1.

To dampen the effect of this wide variability in influent ammonia, dilution water was added to the aeration tank beginning on December 1, 1979. This was done to ensure that ammonia concentrations in the bioreactor would not reach toxic or inhibitory levels during the continuation of start-up. A

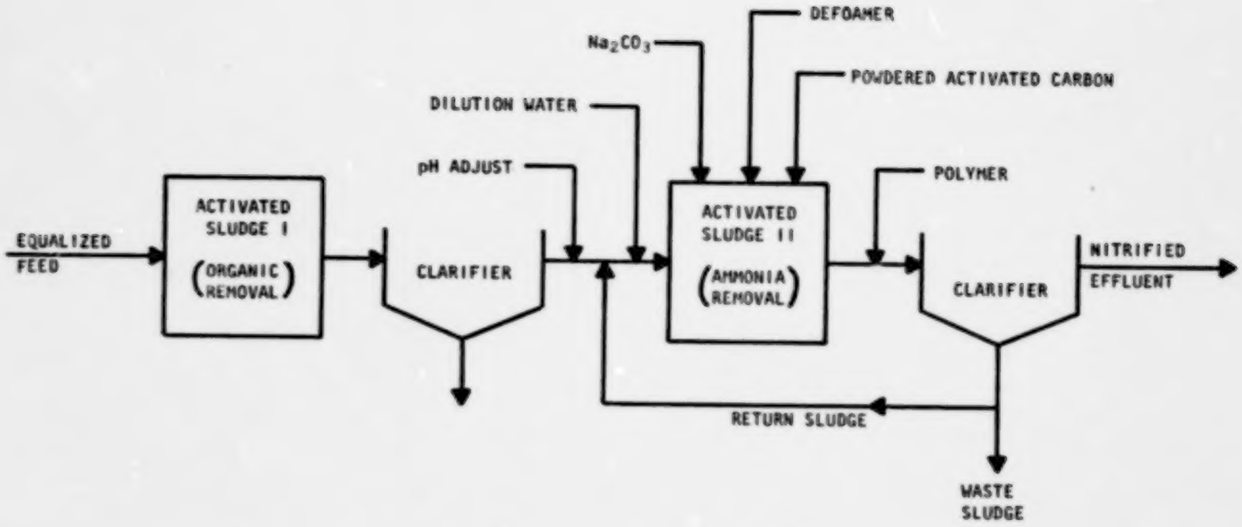

EQUALIZED
FEED
ACTIVATED
SLUDGE I
(ORGANIC REMOVAL)
CLARIFIER
pH ADJUST
DILUTION WATER
Na_2CO_3
DEFOAMER
POWDERED ACTIVATED CARBON
ACTIVATED
SLUDGE II
(AMMONIA REMOVAL)
POLYMER
CLARIFIER
NITRIFIED
EFFLUENT
RETURN SLUDGE
WASTE
SLUDGE

Figure 29. Schematic of activated sludge unit used for nitrification.

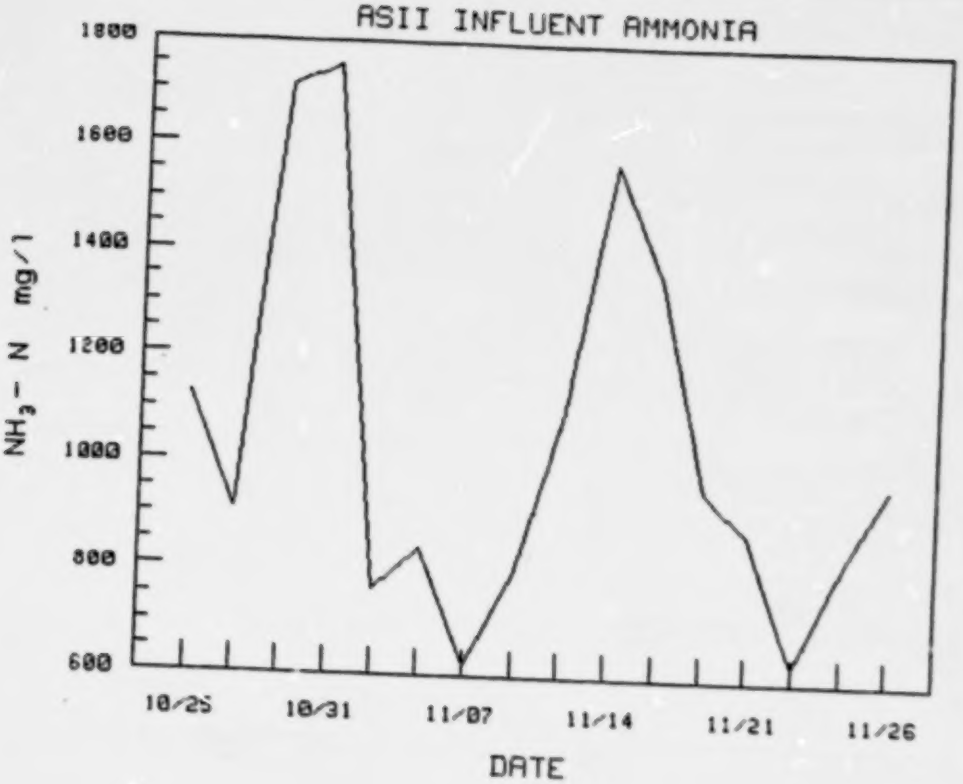


Figure 30. Influent ammonia variability - AS-II (10/22/79 - 11/29/79).

uniform feed ammonia concentration of 200 mg/l was selected for the second stage activated sludge influent. This would be an expected value for a coke plant utilizing well operating ammonia stills. Characteristics of the feedwater with and without dilution are shown in Table 41.

The wastewater exiting the first stage activated sludge unit was analyzed in the field each day for ammonia concentration. A portion of the effluent was wasted and sufficient service water was added to maintain a feed ammonia concentration of about 200 mg/l. As can be seen in Table 41, influent pollutant concentrations were extremely variable. For example, during this period (12/1/79-2/7/80) ammonia varied from 330 mg/l to 977 mg/l. However by diluting the wastewater, a uniform feed concentration could be maintained. Thiocyanate and sulfide were also quite variable without dilution.

TABLE 40. COMPARISON OF INFLUENT VERSUS EFFLUENT VALUES ACROSS AS-II (NITROGENOUS REMOVAL UNIT) OCTOBER 22 - NOVEMBER 29, 1979

Parameter	AS II Influent		AS II Effluent	
	Ave.	Range	Ave.	Range
pH, Units	6.30[a]	5.70-7.90	7.75[a]	7.05-8.20
SS, mg/l	124	42-315	79	43-151
VSS, mg/l	78	39-158	59	31-90
$S^=$, mg/l	3.6	1.6-6.6	1.1	<0.2-1.9
TDS, mg/l	7,690	5,500-9,000	7,850	5,820-8,970
Phenol, mg/l	8.9	0.1-38	0.79	<0.10-2.4
O&G, mg/l	3	<1-5	3	2-4
TKN, mg/l	1,230	874-1,600	1,250	952-1,630
NH_3 as N, mg/l	1,050	620-1,750	1,050	970-1,550
CN_T, mg/l	6.7	2.3-11	6.1	4.2-8.0
CN_A, mg/l	0.29	<0.05-0.80	0.43	<0.05-1.0
TOC, mg/l	301	110-740	119	69-185
TP as P, mg/l	7.1	2.5-16	7.3	3.0-12
BOD, mg/l	475	265-905	97	54-165
Color, APHA Units	1,100	600-1,600	1,000	600-1,500
SCN, mg/l	365	330-400	89	<1-310
Na, mg/l	1,800	-	-	-
K, mg/l	5.0	-	-	-
Ca, mg/l	1.5	-	-	-
Mg, mg/l	5.8	-	-	-
Mn, mg/l	0.4	-	-	-
Fe, mg/l	2.2	-	-	-
Alkalinity as $CaCO_3$, mg/l	104	-	-	-

[a] Typical Value.

TABLE 41. AS-II FEEDWATER CHARACTERISTICS BEFORE AND AFTER DILUTION[a]

Parameter	Feedwater concentration without dilution		Dilution factor		Actual[b] diluted feedwater concentration	
	Ave.	Range	Ave.	Range	Ave.	Range
NH_3, mg/l	567	330-977	0.345	0.109-0.573	160	68-234
CN_T, mg/l	10.5	0.2-25.0			5.0	1.5-13.3
SCN, mg/l	308	2-525			65	<1-180
SS, mg/l	128	48-280			45	10-95
O&G, mg/l	12	2-57			3	<1-7
Phenol, mg/l	8.0	<0.1-90			0.3	<0.1-0.7
Sulfide, mg/l	5.5	0.8-23			0.9	0.1-1.8

a. Average values for period 12/1/79 to 2/7/80.
b. Calculated using dilution factor.

The dilution had a positive effect on the microorganism population and nitrification began to take place in early January, 1980. Excellent ammonia removals were then consistently achieved for the duration of the study. This will be further discussed later in the section.

General Operation Results--
During the acclimation period for the nitrification system, it became evident that several parameters influenced the operation of the activated sludge unit. The most important of these are discussed below in detail.

pH--The optimum pH for nitrification has been reported to be 8.4 (19). Operation at a somewhat lower pH range (7.6 to 7.8) has also been recommended (19). During the pilot study, the pH in the first aeration cell was maintained typically at 7.7, with a range between 6.1 and 8.6 A summary of the pilot study pH profiles for first and third cells of the second stage activated sludge unit as well as the clarifier effluent is shown in Table 42 and presented in detail in Appendix B-3.1.

The pH data for cells 1 and 3 of the three cell aeration tank are plotted in Figure 31. Typically during the study, pH control in the form of alkalinity was required. A dosage of 800 mg/l Na_2Co_3 was typically maintained in the first cell of the aeration tank. This was supplemented with a 25% NaOH dosage of 400 mg/l to help control the pH in the desired range after nitrification began. The caustic feed operated off of a pH controller using a set point of about 7.8.

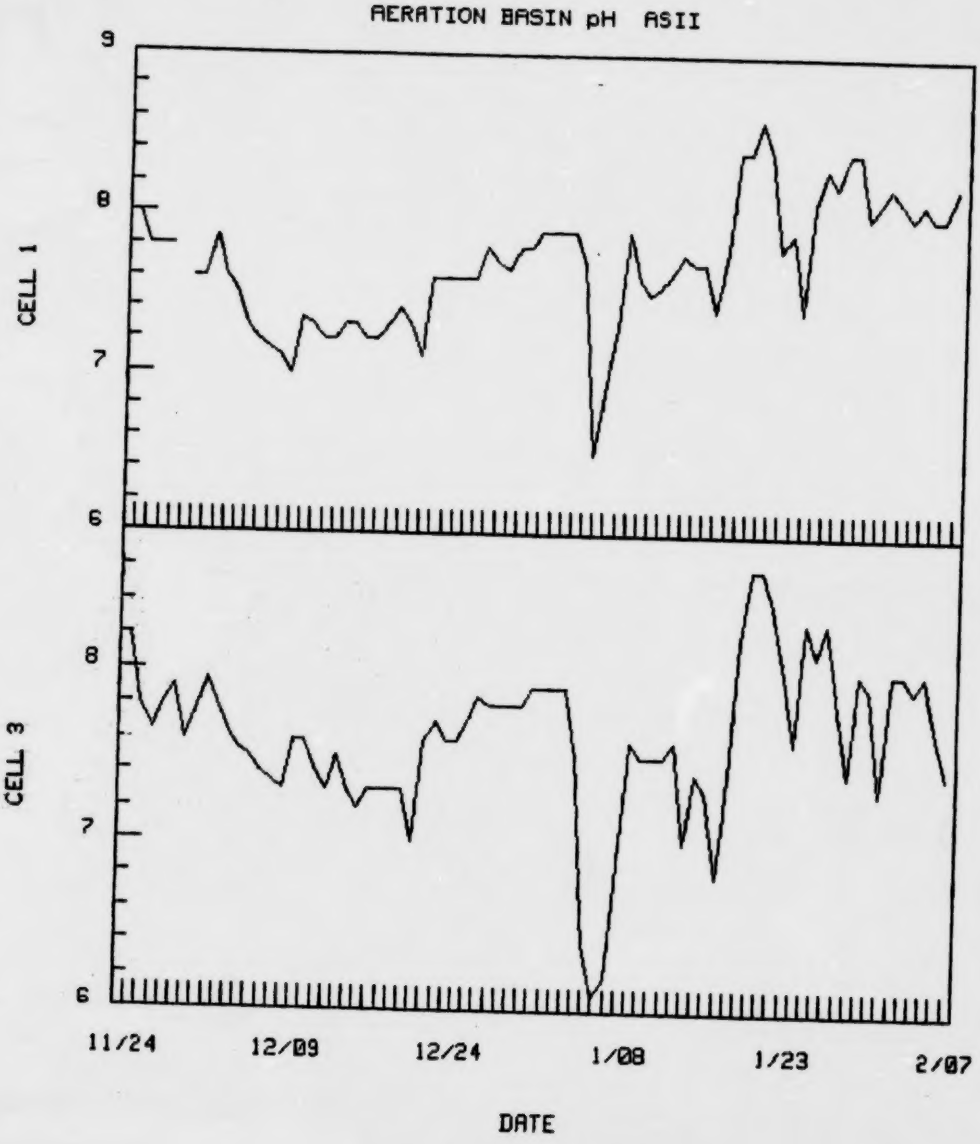


Figure 31. System pH versus date for AS-11.

TABLE 42. pH DATA SUMMARY

Location	Range
Cell 1	6.50-8.60
Cell 3	6.10-8.60
Clarifier Effluent	6.10-8.50

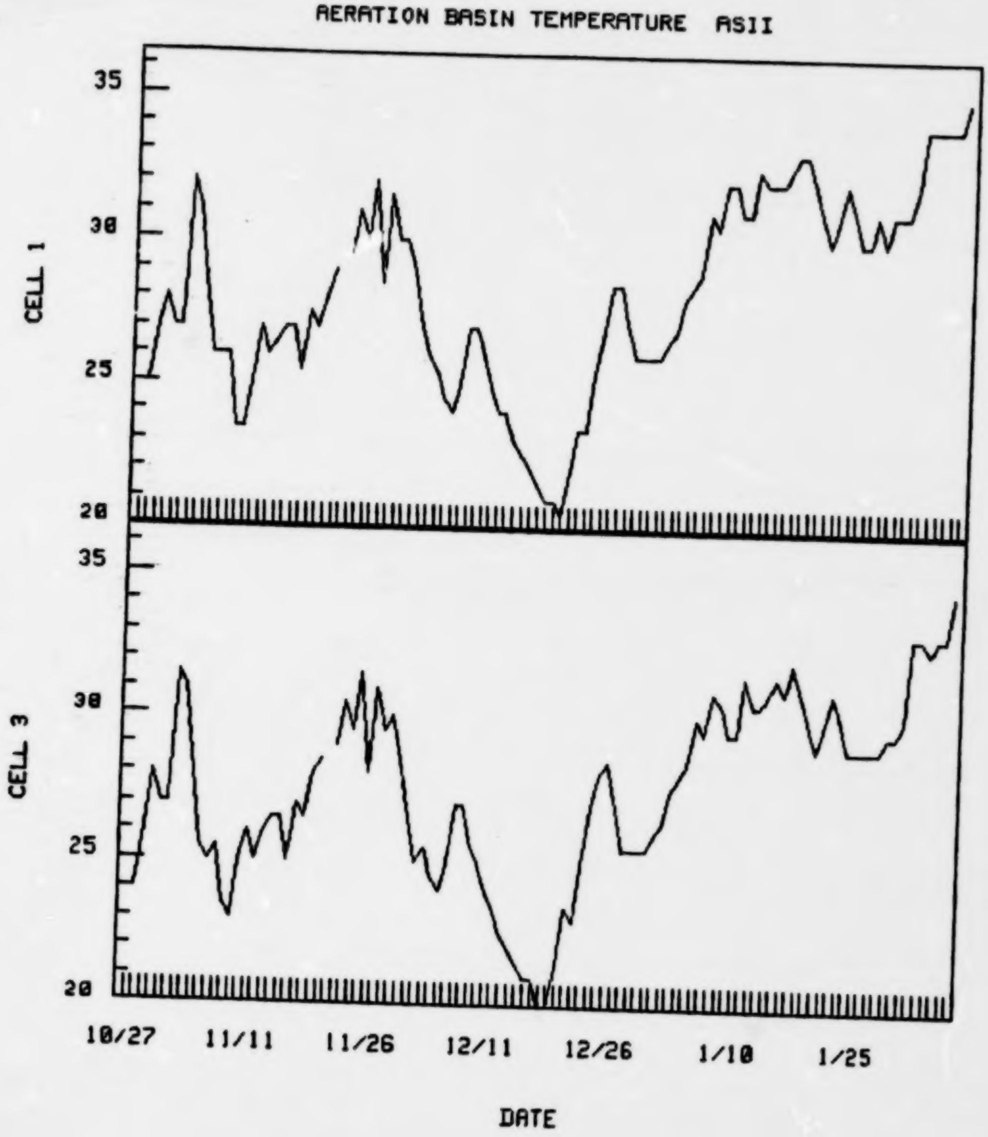
Temperature--Best results in achieving nitrification have been obtained at temperatures between 30 to 35°C (11, 12, 19, 20). During the first seven weeks of the study, wastewater temperature in the first aeration cell averaged 26.5 °C with a range from 21 °C to 32 °C. As winter approached, the wastewater temperature continued to decrease, hitting a low of 20.5 °C on December 18, 1979 as shown in Figure 32.

The mobile bioxidation research trailer was equipped with a temperature control skid and the heater was turned on during the morning of December 18. The heater was installed on the wastewater feed line prior to the first aeration cell of the carbonaceous removal unit (AS-1). The wastewater was heated sufficiently to maintain the second stage system in the optimum operating range. Also all wastewater feed lines were insulated. The increase in temperature in cell 1 of the nitrogenous removal unit is evident from Figure 32. A summary of the entire system temperature profiles is presented in Appendix B-3.2.

The pilot system has a large surface area to volume ratio, and as such, heat loss is great. This would not, however, be the case for a full-scale system. In fact, at the research site investigated, a cooling tower is operated throughout the year to reduce influent temperature to the biological system. Therefore, it is obvious that a temperature control system would not be necessary in any full scale design and no cost allowance has been included.

Foaming--Foaming was experienced in all three cells of the second stage activated sludge aeration tank. Although not as severe as the foaming problem in the first stage unit, a chemical defoamer was required. A typical dosage of 1 mg/l defoamer, added to the first cell of the aeration tank provided adequate control for the entire second stage system.

Dissolved Oxygen--Air was added to the activated sludge system by means of a blower and aeration tank diffuser system provided in the mobile bioxidation trailer. Air was added to each cell of the aeration tank at a flow rate to maintain a dissolved oxygen concentration of about 3 mg/l. Dissolved oxygen readings were taken daily for each cell and a summary of results is presented below in Table 43.



(GRAB SAMPLE DATA)

Figure 32. Aeration basin temperature (°C) for cells 1 and 3 of AS-11.

TABLE 43. DISSOLVED OXYGEN DATA

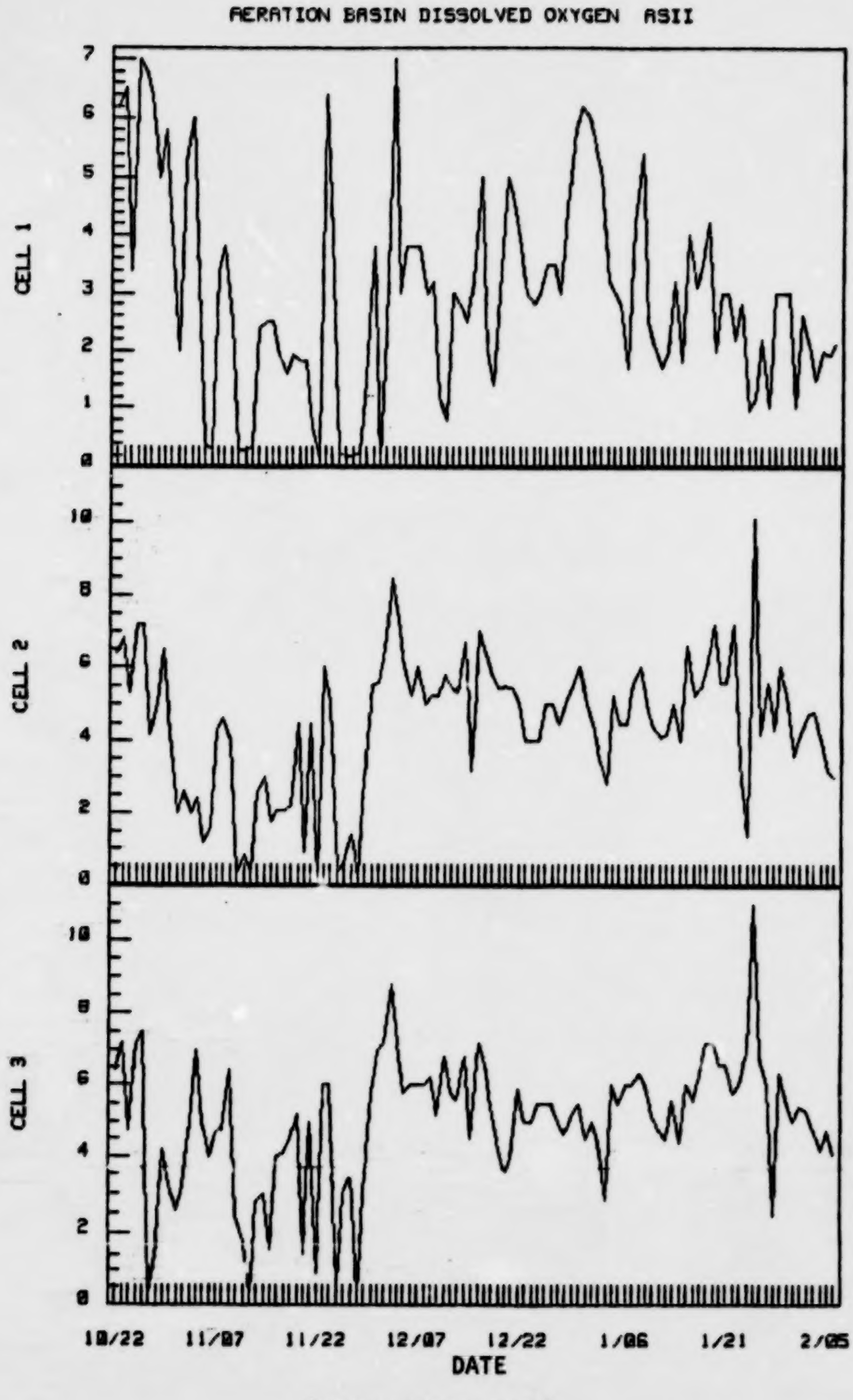
Location	Mean	Std. dev.
Cell 1	3.00	3.07
Cell 2	4.45	3.66
Cell 3	5.00	3.32

As seen in Appendix B-3.3, sufficient dissolved oxygen was routinely maintained until nitrification began to take place in about mid-January, 1980. At this time, dissolved oxygen in the first cell dropped to an average of about 2 mg/l with a range from 0.95 mg/l to 4.2 mg/l. Based on the analytical results however, the infrequent periods of low dissolved oxygen did not have an inhibitory effect on the nitrifying organisms. As shown in Figure 33, cells 2 and 3 had sufficient concentrations of dissolved oxygen throughout the test period.

BOD/Total Kjeldahl Nitrogen (TKN) Ratio--Literature values indicate that for good nitrification, the feedwater to the second stage activated sludge unit BOD to TKN ratio should be maintained between 0.5-1.0 (11). This was used as a guideline in running the nitrification system. As shown in Figure 34, the BOD/TKN ratio varied from a low of 0.25 to a high of 0.7, which was consistently within the optimum range for treatment of ammonia.

Other Factors--During the test work, problems were encountered regarding the settling of sludge (consisting primarily of powdered activated carbon) in a dead spot located in cell 3 of the aeration tank. Once the problem was identified, an air sparger was placed in the dead area to resuspend the sludge. The sparger remained in the tank for the duration of testing and no further problems were encountered. Sludge also had a tendency to settle on the sides of the cone portion of the clarifier resulting in a very dilute return sludge. To correct this problem, it became necessary to manually scrape the sides of the clarifier twice a day. The field operators included this as a regular part of their daily maintenance routine and again no additional problems were observed.

Also during the testing, the mixed liquor was very difficult to settle. As discussed previously, a high density cationic polymer was used. A dosage of 5 mg/l improved effluent clarity, and minimized biomass losses over the weir. During the study, there was no attempt made to optimize the polymer dosage. However, screening tests were performed to select the optimum type of polymer.



(GRAB SAMPLE DATA)

Figure 33. System dissolved oxvgen (mg/l) versus date for AS-11.

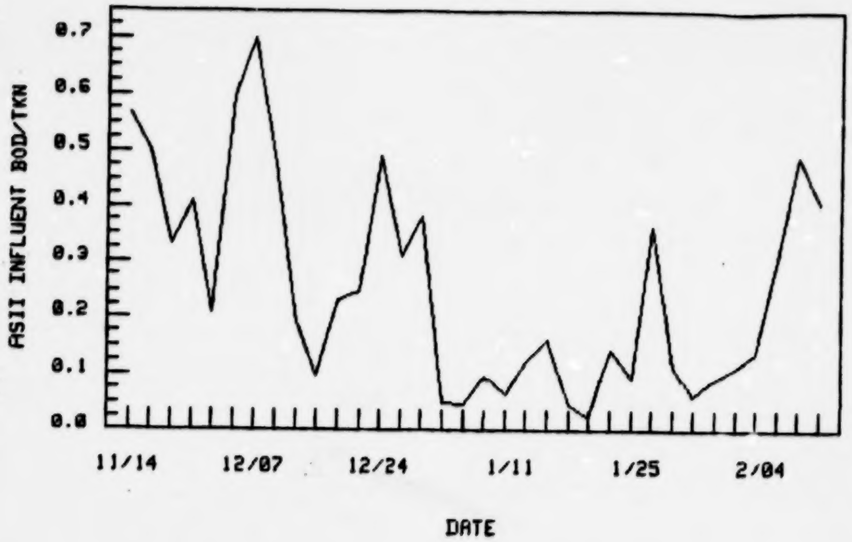


Figure 34. BOD/TKN ratio for the feedwater to AS-II.

Other Treatment Results--In addition to the ammonia and cyanide results discussed previously, it is interesting to examine the results of treatment for additional parameters. For example, during the period of active nitrification (January 6 - February 7, 1980), TKN was reduced from 183 mg/l (calculated concentration of the feedwater after dilution) to 29 mg/l, a reduction of 86 percent. There was no noticeable reduction in oil and grease or phenol, however, these parameters were already reduced to low levels by the first stage activated sludge system (carbonaceous removal). Interferences developed in the standard sulfide test once nitrification took place for unknown reasons; however, prior to nitrification, influent sulfide to the second stage activated system was typically less than 0.2 mg/l. During nitrification, the second stage unit was also effective in removing the residual thiocyanate that was present in the AS-II feedwater. During the period between January 6, 1980 and February 7, 1980, influent thiocyanate to the second stage unit average 253 mg/l prior to dilution.

After dilution, this concentration was reduced to 65 mg/l. AS-II effluent thiocyanate concentration was consistently less than 1 mg/l.

Figure 35 illustrates the AS-II effluent suspended solids pattern over time. Prior to diluting the feedwater (before December 1, 1979) suspended solids averaged 124 mg/l. The effect of dilution is evident by the steep drop in effluent suspended solids seen in Figure 35 at the end of November. During the period January 9 - February 7, 1980, AS-II diluted feed suspended solids averaged 45 mg/l. Effluent suspended solids were also typically less than 50 mg/l until January 23, 1980 when they began to increase as a result of discontinuing the use of polymer. The polymer addition was stopped during the intensive test period at the end of the study so that priority pollutant analyses would not be influenced by the chemical usage. The rise of effluent suspended solids illustrates the need for polymer to improve settling and reduce loss of biomass over the effluent weir.

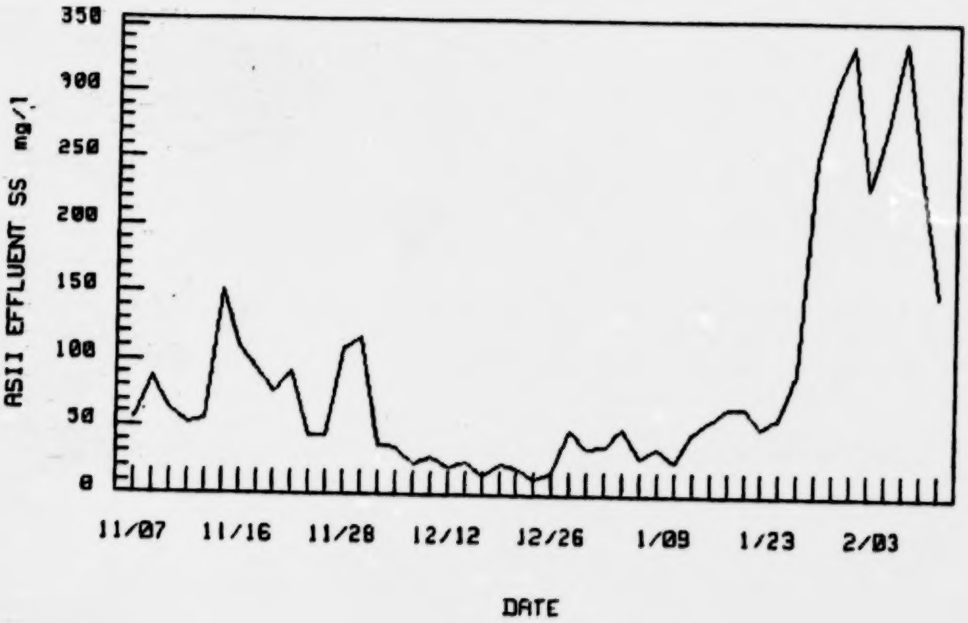


Figure 35. AS-II effluent suspended solids with time.

Operation During Nitrification Phase--

Once a wastewater with ammonia levels low enough to avoid interference with the treatment process was provided to the second stage activated sludge unit through dilution (December 1, 1979 - Feburary 7, 1980), the nitrifying bacteria population continued to increase and evidence of nitrification was observed in early January, 1980. Excellent ammonia removals were then consistently achieved for the duration of the project as shown in Figure 36. The graph shows that a period of acclimation from December 1, 1979 to January 9, 1980 was required to obtain optimum ammonia removal. Effluent ammonia was initially higher than influent until the entire system was brought into equilibrium with the dilution water. A summary of the influent and effluent pollutant concentrations across the second stage activated sludge unit during the period of active nitrification (January 9 - February 7, 1980) is shown in Table 44. During this period, ammonia removal averaged 90.6 percent. There was also some removal of thiocyanate and sulfide. The dilution factors used for the AS-II feedwater are shown in Appendix B-4.1. A summary of the daily analytical values for the AS-II influent and the second stage clarifier effluent is shown in Appendix B-1. Notice from Figure 37 that a decrease in effluent ammonia resulted in an increase in effluent nitrite-nitrate. During the nitrification period, effluent nitrite-nitrate averaged 186 mg/l.

Reviewing Appendix B-1, it is evident that once nitrification took place, total cyanide effluent values began to increase and actually exceeded the influent concentrations. This has been reported in the literature as "Nitrite Interference". It has been previously shown that the standard test for cyanide gave erroneously high values when approximately 20 mg/l NO_2-N was present along with nitrite-reactive compounds (22). They react to form hydrogen cyanide (HCN) under the acid conditions present during the distillation step. Various researchers have reported these nitrite-reactive substances include aliphatic and aromatic amines, phenol, aniline, methyl isobutyl ketone and others. The presence of nitrite-reactive organic compounds would be expected in coke plant wastewater.

The formation of hydrogen cyanide (HCN) during distillation via nitrosation of nitrite-reactive compounds would probably only occur in the second stage activated sludge effluent after the nitrification process started about January 7, 1980. Sufficient nitrite levels probably were not present previous to nitrification. After nitrification began in the second stage activated sludge system, nitrite-nitrate (NO_2/NO_3) levels increased from an average of 1-2 mg/l to 160-215 mg/l. Concentrations ranging from <1-2 mg/l NO_2/NO_3-N in the system feedwater and the first stage activated sludge effluent apparently were not sufficient to produce a significant amount of cyanide.

It has been shown that cyanide levels increase with the addition of nitrite eventually reaching a limit probably due to depletion of nitrite-reactive compounds (21). It would be difficult to make a correction for cyanide based nitrite analysis since various wastewaters respond differently to a given nitrite level.

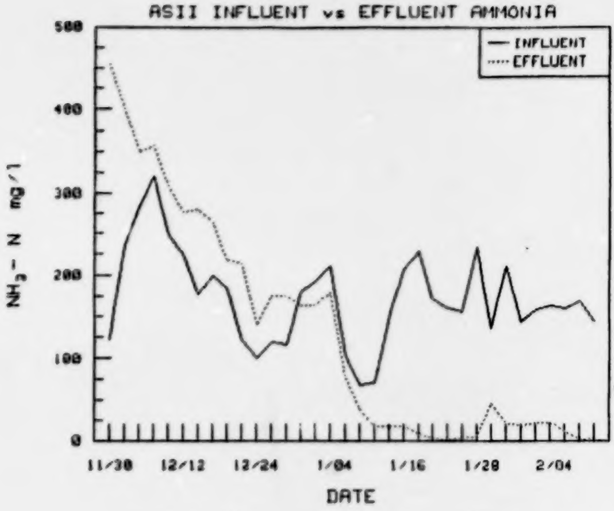


Figure 36. Influent and effluent ammonia values for a selected period of AS-II operation (12/1/79-2/7/80).

TABLE 44. COMPARISON OF INFLUENT (WITH DILUTION) VERSUS EFFLUENT VALUES ACROSS AS II (NITROGENOUS REMOVAL UNIT) JANUARY 9 - FEBRUARY 7, 1980

Parameter	AS II Influent		AS II Effluent	
	Ave.	Range	Ave.	Range
SS, mg/l	45	10-95	155	25-340
VSS, mg/l	25	6-51	93	13-230
S, mg/l	0.9	0.1-1.8	I	-
Phenol, mg/l	0.3	<0.1-0.7	0.3	<0.1-0.9
O&G, mg/l	3	<1-7	9	3-21
TKN, mg/l	183	78-267	25	8-73
NH_3 as N, mg/l	160	68-234	15	1-45
CN_T, mg/l	5.0	1.5-13.3	8.2	2.1-30
CN_A, mg/l	0.49	<0.05-3.1	0.57	<0.05-5.1
TOC, mg/l	68	28-190	73	21-158
TP as P, mg/l	2.7	1.0-4.1	2.9	0.9-4.6
BOD, mg/l	30	4-97	25	4-87
Color, APHA Units	300	150-600	400	100-600
SCN, mg/l	65	<1-180	5	<1-65
NO_2/NO_3 as N, mg/l	<1	-	186	160-215

I: Denotes Interference

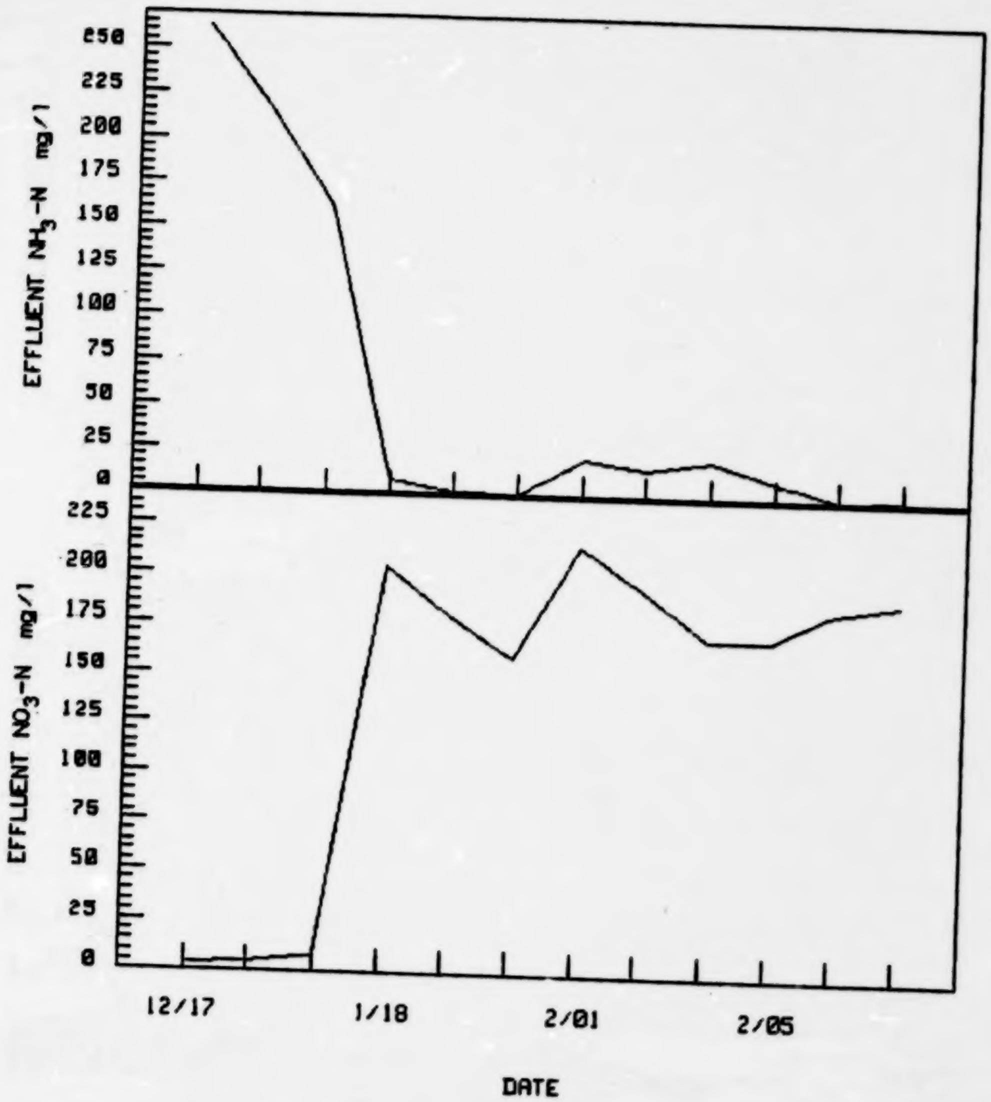


Figure 37. Effluent ammonia and nitrite/nitrate values during period of active nitrification - AS-II.

Table 45 summarizes effluent pollutant concentrations from AS-II during the period of active nitrification.

TABLE 45. AS-II EFFLUENT CHARACTERISTICS FOR BIOLOGICAL RESEARCH SITE

Parameter	Pilot study effluent value, mg/l[a]
CN_T	8.2
Phenol	0.3
NH	15
O&G	9
SS	155
Sulfide	I
SCN	5

a. For period of testing 1/6/80-2/7/80

I - Denotes interference.

From the data listed above, it is obvious that effluent ammonia levels were quite low. The problem associated with total cyanide analysis has already been discussed; however, it should be noted that cyanide-A values were typically <0.05 mg/l. Similarly, phenol values were mostly <0.1 mg/l, with the exception of a few high values. Oil and grease effluent levels were also consistently low. The suspended solids pilot value shown in Table 45 (155 mg/l) includes a period in which polymer was not being added to the final clarifier (1/30-2/7/80). Prior to this time, suspended solids values were consistently less than 50 mg/l. However, polymer addition would be required. As discussed previously, during the nitrification phase interferences developed in the sulfide test, which prevented the obtaining of any data for this parameter. Prior to nitrification, sulfide analytical values of <0.2 mg/l were commonly achieved. With regard to thiocyanate, all reported values were 1 mg/l with the exception of one (65 mg/l). This increased the overall average to 5 mg/l.

Sludge volume index (SVI) throughout the study was typically less than 60 and only exceeded 100 on five occasions. For activated sludge, a SVI of 100 or less is indicative of a well settling sludge. Sludge settling rate was also measured daily as a part of the field operator's daily work schedule. Settling rates ranged from 0.002 m/min (0.008 ft/min) to

0.14 m/min (0.47 ft/min). The average mixed liquor settling rate was 0.043 m/min (0.14 ft/min). Complete information on SVI and settling rate daily values determined during the study can be found in Appendix B-4.2. Information on mixed liquor suspended solids, mixed liquor volatile suspended solids, return sludge solids and return sludge volatile suspended solids can also be found in the Appendix B-4.2. The increase in MLSS and MLVSS with time can be seen in Figure 38. Once dilution was started (December 1, 1979), a uniform feed ammonia concentration was maintained and the biomass went into an active growth phase. Sufficient nitrifiers were eventually present in the activated sludge to remove the influent ammonia to acceptable levels. MLSS values ranging from 6,000 mg/l - 9,000 mg/l were common during the nitrification phase. Of this, approximately 85 percent of the MLSS was volatile. Sludge volume index is also plotted in Figure 38.

Food to Microorganism Ratio (F/M)--The food to microorganism ratio (F/M) for the nitrification system as discussed below is defined as:

$$F/M = \text{kg ammonia/day/kg MLSS}$$

Successful operation in treating coke plant wastewater at an ammonia loading of 0.02 lb ammonia/day/kg MLSS has been previously reported in the literature (22). During the start-up and stabilization of the nitrification system, the F/M ratio varied from 0.008 day^{-1} to 0.06 day^{-1} (11/30/79-1/11/80). However, once there was an active growth of biomass and nitrification was occurring, the F/M ratio stabilized for the remainder of the study. This would be the period between January 14 and February 7, 1980, and is illustrated in Figure 39. During this latter period, the F/M ratio averaged 0.023 day^{-1}, and ranged between 0.017 day^{-1} and 0.028 day^{-1}. A complete listing of the daily F/M ratios investigated is presented in Appendix B-4.3.

Appendix B-4.3 also summarizes the results of ammonia removal achieved, the daily MLSS concentrations that the nitrification system was operated at, and the hydraulic retention times investigated. During the study, hydraulic retention time remained stable, varying only from 24.8 hours to 29.9 hours. Since nitrification did not take place until late in the pilot study, no data were generated on sludge age.

For purposes of design, three operating periods were selected during the study which had stable F/M ratios and active nitrification. The first selected period (1/9/80-1/14/80) had an average F/M ratio of 0.014 day^{-1}. During period two (1/16/80-1/23/80), an average F/M ratio of 0.024 day^{-1} was maintained. The final selected period (1/30/80-2/7/80) had an F/M ratio which averaged 0.025 day^{-1}. Table 46 summarizes the periods selected for design.

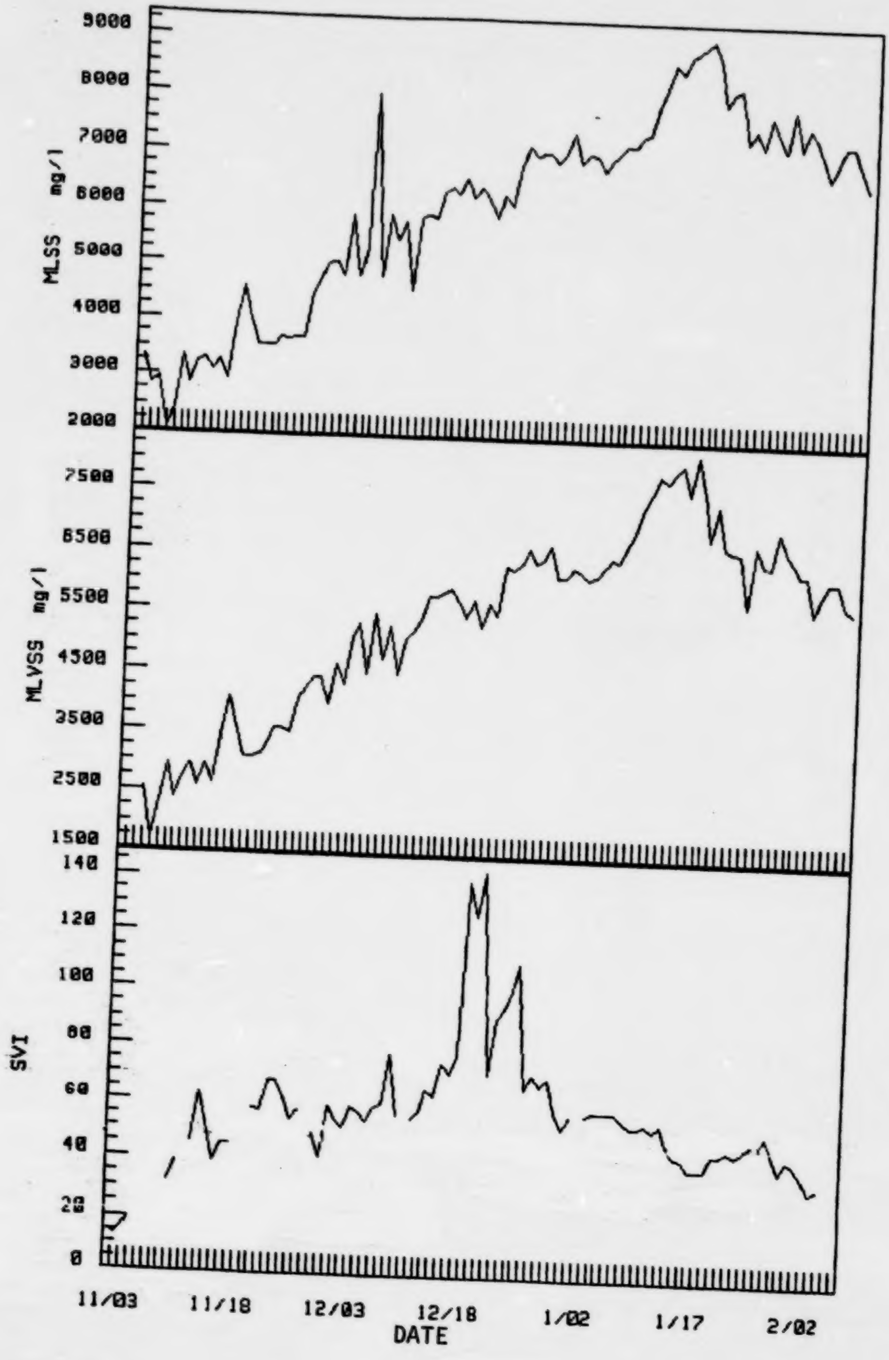


Figure 38. Plots of MLSS, MLVSS and SVI versus time - AS-II.

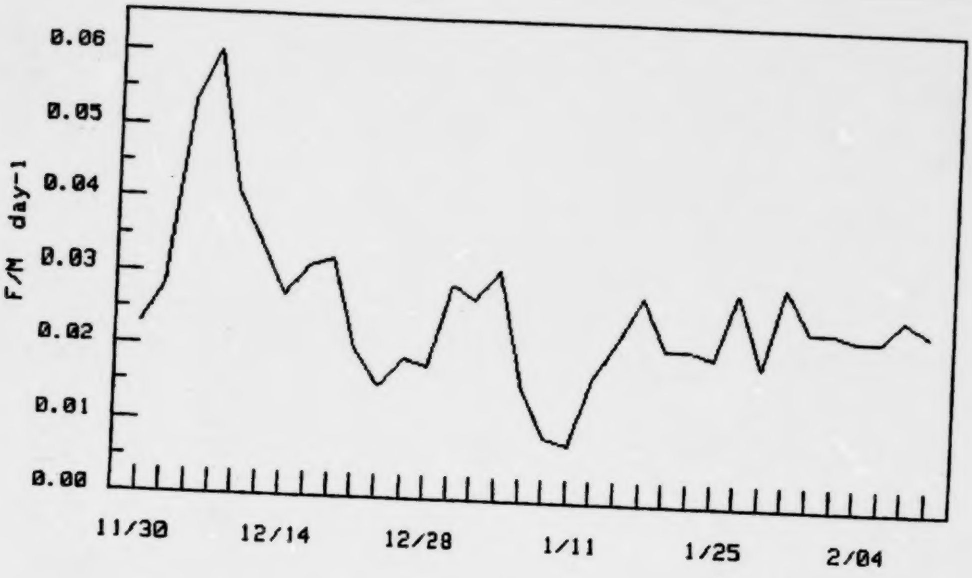


Figure 39. F/M ratio versus time - AS-II.

TABLE 46. THREE SELECTED PERIODS OF CONSISTENT OPERATION FOR AS-II

Period	Ave. F/M day^{-1}	Ave. NH_3 removal, %	Ave. SVI	Ave. settling rate, ft/min
1/9-1/14/80	0.014	78.8	55	0.10
1/16-23/80	0.024	96.6	43	0.16
1/30-2/7/80	0.025	89.2	40	0.22

Note: ft/min x 0.3048 = m/min

As seen in Table 46, optimum ammonia removal occurred at an F/M ratio of 0.024 day^{-1} (96.6 percent ammonia removal). At a slightly higher F/M ratio (0.025 day^{-1}), ammonia removal dropped slightly, but improvements were seen in the settling rate and Sludge Volume Index (SVI). Complete data for these selected test periods are presented in Appendices B-4.4 and B-4.5.

Special Testing--Similar to the special testing performed on unit AS-I, oxygen uptake rates and gravity thickening tests were conducted on unit AS-II. The test work was undertaken late in the investigation at the same time the experimental studies were being run on unit AS-I. During this period, unit AS-II was operating at steady state.

For the oxygen uptake tests, raw waste and mixed liquor were blended in pre-selected ratios to determine oxygen uptake rates at different F/M ratios. The results of these tests are shown below in Table 47.

TABLE 47. OXYGEN DEMAND AT VARIOUS F/M RATIOS - AS-II

Date	Time	Location	MLVSS (mg/l)	Influent NH_3-N, (mg/l)	F/M ratio (kg NH_3-N/day/ kg (MLVSS)	O_2 Demand (mg/l/hr)
2/5/80	1430	AS-II	6056	52	0.008	53
2/5/80	1500	AS-II	6062	103	0.016	69
2/5/80	1530	AS-II	6069	153	0.026	77
2/5/80	1600	AS-II	6075	204	0.031	85
2/7/80	1430	AS-II	5338	53	0.009	62
2/7/80	1500	AS-II	5345	107	0.018	77
2/7/80	1530	AS-II	5353	160	0.027	98
2/7/80	1600	AS-II	4342	174	0.037	124

It is evident from examining the Table 47 results that oxygen demand increased with an increase in F/M ratio. The uptake rate is the amount of oxygen that would have to be supplied to an activated sludge unit to maintain microorganism metabolism.

Each aeration cell of the pilot system was checked for oxygen uptake using a respirometer. The results of testing are shown in Appendix B-3.10. Oxygen requirements were highest in the first cell and decreased as the waste passed through the system to cells 2 and 3. The information obtained in the oxygen uptake study was used to estimate air requirements for the full-scale system design which is presented later in this section.

During the intensive test period, gravity thickening tests were also conducted on the nitrified activated sludge. Two return sludges were investigated, 2/2/80 and 2/6/80. Acceptable dilution ratios were selected and the corresponding settling rates were recorded as shown below in Table 48. Using this information, flux concentration curves were prepared and these illustrations are presented in Appendices B-4.6 and B-4.7.

TABLE 48. SETTLING DATA FOR FLUX CONCENTRATION CURVE

Source	Date	Sludge conc. (mg/l SS)	Settling rate, m/hr
Return Sludge-II	2/2/80	14,800	0.0544
Return Sludge-II	2/2/80	17,00	0.030
Return Sludge-II	2/2/80	18,400	0.0212
Return Sludge-II	2/2/80	21,000	0.0155
Return Sludge-II	2/2/80	23,700	0.013
Return Sludge-II	2/2/80	27,100	0.011
Return Sludge-II	2/2/80	29,700	0.010
Return Sludge-II	2/6/80	17,000	0.068
Return Sludge-II	2/6/80	19,000	0.0435
Return Sludge-II	2/6/80	21,000	0.0295
Return Sludge-II	2/6/80	23,000	0.021
Return Sludge-II	2/6/80	25,000	0.017
Return Sludge-II	2/6/80	27,000	0.0145
Return Sludge-II	2/6/80	31,000	0.012
Return Sludge-II	2/6/80	36,000	0.010
Return Sludge-II	2/6/80	48,000	0.0073

Using Appendices B-4.6 and B-4.7 as a guide, it was concluded that a full-scale gravity thickener could obtain an underflow suspended solids concentration of 32,000 mg/l. The corresponding mass loading would be 21.6 kg/m^2/day (4.42 lb/ft^2/day). Increasing the thickener underflow concentration would be possible, but would result in a reduced mass loading as shown in Table 49.

Full-Scale System Design, Costs, and Space Requirements

Introduction--

The design of a full-scale nitrification system (Stage II) depends upon fewer process parameters than for a carbonaceous system (Stage I). The system sizes, and also costs, depend mainly on ammonia concentration and mixed liquor volatile suspended solids (MLVSS) concentration.

Secondary considerations, such as thiocyanate concentration, were of less

TABLE 49. MASS LOADINGS AT VARIOUS SLUDGE UNDERFLOW CONCENTRATIONS

Source	Date	Thickener underflow conc. (mg/l SS)	Maximum mass loading kg/day/m^2
Return Sludge-II	2/2/80	24,000	39.4
Return Sludge-II	2/2/80	28,000	27.2
Return Sludge-II	2/2/80	32,000	21.6
Return Sludge-II	2/2/80	36,000	18.4
Return Sludge-II	2/2/80	40,000	16.2
Return Sludge-II	2/6/80	32,000	41.0
Return Sludge-II	2/6/80	36,000	32.0
Return Sludge-II	2/6/80	40,000	26.8
Return Sludge-II	2/6/80	44,000	23.6
Return Sludge-II	2/6/80	48,000	21.2

importance since treatment in Stage I reduced the effect of these contaminants. Furthermore, the addition of powdered activated carbon (PAC) to Stage II reduced the toxic effects of wastewater contaminants and other constituents on the nitrifiers.

The nitrifiers pilot system was operated under stable conditions to generate design data. Bench tests were performed to develop settling rates and oxygen uptake rates when it became apparent that nitrification was occurring in the pilot system. Since the pilot system was operating at design conditions, bench test data would be directly applicable for design purposes.

The MLSS bar graph (Figure 38) for the nitrification pilot system was compared to the field log book entries. Operating periods were selected on the basis of increasing MLSS, which indicated healthy nitrifier growth. Other factors, such as ammonia removal efficiency, good clarifier operation, low SVI, etc. were also considered. The same criteria were used for selection of Stage II design periods as for selection of Stage I design periods, as discussed in a previous section of this report.

Two wastewater treatment systems were designed. Design No. 1 is based on pilot plant data and the condition of wastewater at the test site. Design No. 2 assumes a feedwater from a well operating ammonia still. Characteristics of each type of wastewater are shown in Table 50.

Choice of Steady State Periods for Design and Characterization of Performance--The second stage of the pilot system was operated under three distinct sets of conditions. Appendix B-4.4 shows the data from these three periods. Time constraints prevented operation of Stage II under a wider range of conditions, however, the data were adequate for the development of design factors. The average Stage II operating conditions are shown in Table 51.

TABLE 50. STAGE II FEEDWATER CHARACTERISTICS USED FOR THE DESIGN OF THE NITRIFICATION SYSTEMS

Parameter	Design No. 1 waste as received	Design No. 2 assumes waste from a well-operating ammonia still
Design influent BOD, mg/l	200	100
Design influent NH_3-N, mg/l	700	200
Design effluent NH_3-N, mg/l	32	32
Design effluent SS, mg/l	86	86
Design settling rate, ft/min	0.10	0.10
Dilution ratio	3.5:1	None
--		

NOTE: ft/min x 0.3048 = m/min

TABLE 51. AVERAGE OPERATING CONDITIONS FOR STAGE II DURING THREE SELECTED PERIODS

Parameter	Test Period		
	1/9-1/14/80	1/16-1/23/80	1/30-2/7/80
Influent ammonia, mg/l[a]	113	204	167
Effluent ammonia, mg/l	24	7	18
MLSS, mg/l	7,189	8,038	7,100
MLVSS, mg/l	6,396	5,908	5,189
Detention time, days	1.09	1.19	1.09
Flow, ℓ/min[a]	3.8	3.5	3.78
Effluent SS, mg/l	35	60	266[b]

a. Includes dilution water.
b. No polymer was used.

Stage II Design Criteria--

Selection of the design MLVSS concentration was based on pilot plant operation. One investigator showed that by increasing MLVSS from 2,500 mg/l to 3,500 mg/l, ammonia removal improved by 25 percent (16). Increasing MLVSS above that level should further improve ammonia removal. By selecting a design MLVSS concentration of 7,000 mg/l, the MLVSS concentration would be in the middle of the optimum reported range (11) and consistent with the successful performance of the pilot plant. Also, another investigator reported operation at a MLVSS concentration of 8,000 mg/l (22).

The ratio of effluent ammonia concentration to influent concentration is the other important design parameter (16). The method outlined in this reference (16) was used to size the aeration basin of the nitrification system.

Influent ammonia concentration was assumed to remain constant as the wastewater flows through Stage I. Theoretically, some of the ammonia would be oxidized in Stage I to satisfy the C:N:P requirement. However one investigator reported that thiocyanate oxidizes to form ammonia (23). Thus, the uptake of ammonia as nitrogen is offset by the production of ammonia by the thiocyanate. The pilot plant data verified that the net change in ammonia was zero as the flow passed through Stage I.

The ammonia concentration in the nitrification system influent has a great effect on design. Coke plant effluent may contain ammonia concentrations as high as 2,000 mg/l (11). Investigators have found that the upper limits of ammonia concentration are 1,600 mg/l (24) to 2,000 mg/l (12). One investigator recommends 600 mg/l as the upper limit for ammonia concentration (23). Coke plants having ammonia stills typically discharge wastewater having ammonia concentrations of 200 mg/l (11). Other coke plants must dilute their wastewater to reduce their ammonia concentration to 200 mg/l, The full-scale design was based on the 200 mg/l ammonia concentration as previously recommended (11) and verified by the pilot scale tests.

The literature reports sludge age may range from 6-80 days (3, 6, 22). One investigator recommends a sludge age of more than 6 days (16) for 80 percent nitrification, while another recommends 30 days (14). The investigator who used PAC recommended a sludge age of 7-12 days (3). Longer sludge age has the effect of increasing PAC residence time in the system, which helps to reduce the effects of shock loadings. Since the nitrification system in this design would use PAC, the sludge age could be between 7-12 days, as recommended, or up to 30 days.

The concentration of PAC in the aeration tank was maintained at approximately 950 mg/l during the pilot plant tests. The ratio of PAC:MLSS must be considered when wasting sludge, as the rate of applying make-up PAC will depend upon the amount lost. The amount of PAC lost as effluent suspended solids was neglected in the design calculations.

Since nitrification produces light, fluffy flocs, polymer must be added to aid clarification. The pilot system data show that effluent suspended solids increased drastically when polymer was not added. A dose of 5 mg/l of polymer was assumed for design and cost purposes.

Since foam was a problem during the pilot plant tests, it is recommended that defoamer be used on full-scale designs. Water sprays are useful, and defoamer addition at 1-2 mg/l should be sufficient (12). A liquid defoamer dose of 2 mg/l was used for design costs.

During the nitrification process, 7.14 mg of alkalinity as $CaCO_3$ is destroyed per milligram of ammonia nitrogen oxidized (18). During the pilot study, sodium carbonate (Na_2CO_3) was added in an amount sufficient to provide an effluent alkalinity of about 100 mg/l as $CaCO_3$. To maintain the pH in the optimum range, caustic soda was also used to supplement the sodium carbonate. Quantities of alkalinity supplement used in the full-scale designs are shown in Tables 52 and 53.

The design ammonia removal rate is within the range of ammonia removals experienced with the pilot size system. Therefore, it is reasonable to assume that a full-scale nitrification system will remove similar amounts of ammonia when operated under similar conditions. The design parameters, as selected on the basis of the literature and as computed from the pilot scale data, are tabulated in Table 54.

Equipment Description, Costs, and Space--
The aeration basin and clarifier for Design No. 1 will be larger due to the addition of dilution water. Chemical use and power costs will also be greater. No cost is included for dilution water, since supplies of non-contact cooling water, river water, or other sources of supply are usually available at little or no cost.

Equipment Description, (Stage II, Design No. 1)--The nitrification system, Stage II, would consist of aeration tankage, clarification, sludge recycle system, blowers, and chemical feed systems. This equipment is shown by the diagram in Figure 40. Equipment sizes, quantity, and other descriptive information are shown in Table 52. Equipment costs are given in Table 53.

The aeration tanks would be built of cast-in-place concrete below ground. Each aeration tank would be fitted with fiberglass pipes with coarse bubble diffusers for air distribution. Air would be supplied to the aeration tanks by blowers. Table 52 lists the total air flow required for each size system. The number of blowers listed includes a spare blower which does not contribute to the total air supplied unless one of the other blowers fails.

Clarifiers would be built of cast-in-place concrete below ground. The clarifiers would be of the type having peripheral effluent weirs and rotating sludge collection manifolds.

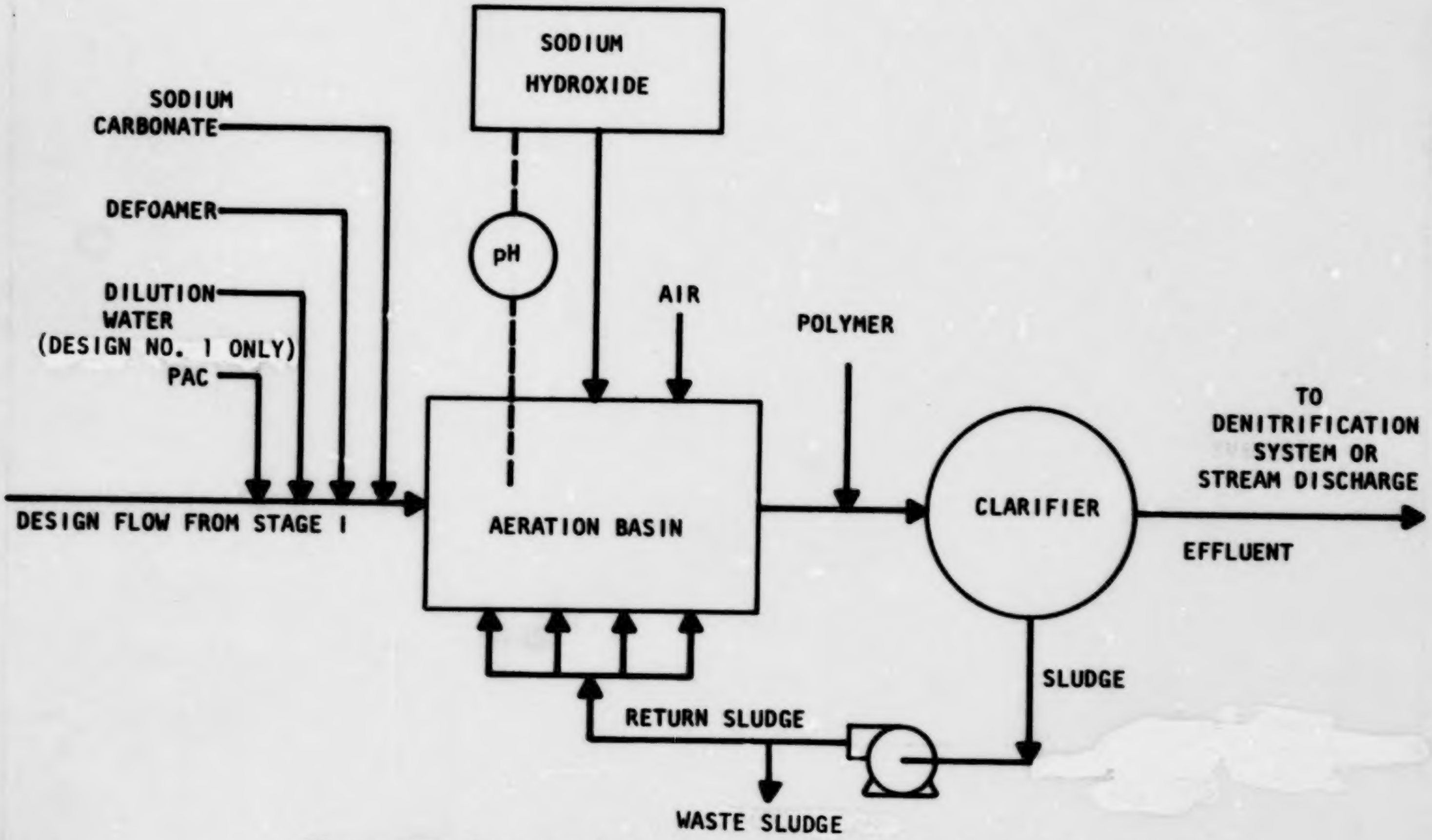

SODIUM HYDROXIDE
SODIUM CARBONATE
DEFOAMER
DILUTION WATER (DESIGN NO. 1 ONLY)
PAC
pH
AIR
POLYMER
DESIGN FLOW FROM STAGE I
AERATION BASIN
CLARIFIER
TO DENITRIFICATION SYSTEM OR STREAM DISCHARGE
EFFLUENT
SLUDGE
RETURN SLUDGE
WASTE SLUDGE

Figure 40. Diagram of Stage II completely mixed nitrification system for coke plant wastewater (Design No. 1 and Design No. 2).

TABLE 52. EQUIPMENT DESCRIPTION - STAGE II - DESIGN NO. 1

Flow rate, m^3/day (mgd)	379 (0.1)	1892 (0.5)	3785 (1.0)
Total aeration volume, m^3 (ft^3)	1,930 (68,200)	9,630 (340,000)	19,300 (680,000)
Aeration tank number	2	3	5
Aeration tank size (WxLxH) m (ft)	9x23x5 (30x75x15)	9x77x5 (30x252x15)	9x92x5 (30x303x15)
Total air required, m^3/min (cfm)	42 (1480)	210 (7400)	420 (14,800)
Blower number	3	3	3
Defoamer[a], l/day (gpd)	2.6 (0.7)	13 (3.5)	30 (8.0)
Sodium hydroxide[b], l/hr (gph)	23 (6)	110 (30)	230 (60)
Powdered carbon[c], l/hr (gph)	19 (5)	95 (25)	200 (50)
Sodium carbonate[d], l/hr (gph)	946 (250)	4730 (1150)	9460 (2500)
Polymer[e], l/hr (gph)	300 (80)	1460 (385)	3000 (800)
Total clarifier area, m^2 (ft^2)	60 (650)	300 (3240)	600 (6480)
Clarifier number	1	2	2
Clarifier diameter, m (ft)	9.1 (30)	13.7 (45)	19.8 (65)

a. Based on 2 mg/l dosed.
b. Based on 50% concentration.
c. Based on a 143 mg/l dosed and 0.12 kg/l stock.
d. Based on 1840 mg/l dosed and 0.12 kg/l stock.
e. Based on 5 mg/l dosed and a stock solution of 1 mg/ml.

NOTE: Because of a 3.5:1 dilution ratio, actual flow through the system is 3.5 times the feed flow rate listed.

TABLE 53. COST SUMMARY - STAGE II DESIGN NO. 1

Design flow, m^3/day (mgd)	379 (0.1)	1892 (0.5)	3875 (1.0)
Capital costs, \$[a]			
Wastewater pumps	7,000	10,000	14,000
Chemical pumps	15,000	19,000	21,000
Chemical mixers	3,000	5,000	8,000
Chemical tanks	12,000	32,000	52,000
Aeration tanks	69,000	274,000	561,000
Aeration piping	19,000	120,000	243,000
Aeration blowers	25,000	73,000	136,000
Clarifier	56,000	154,000	210,000
pH controller	4,000	4,000	4,000
Electrical panel	8,000	17,000	26,000
Delivered equipment total	218,000	708,000	1,275,000
Equipment installation	87,000	283,000	510,000
Installed equipment total	305,000	991,000	1,785,000
Process piping	92,000	297,000	536,000
Instrumentation	15,000	49,000	89,000
Capital equipment total	412,000	1,337,000	2,410,000
Facilities & fees	557,000	1,872,000	3,374,000
Capital cost total	989,000	3,209,000	5,784,000
Operating cost, \$			
Electrical	21,000	87,000	173,000
Operator	36,000	36,000	36,000
Maintenance labor & materials	21,000	67,000	120,000
Chemicals	418,000	2,091,000	4,183,000
Yearly operating subtotal	496,000	2,281,000	4,512,000
Capital Amortization			
20 yr life @ 7.125%	94,000	306,000	551,000
Total Annualized cost, \$/yr[b]	590,000	2,587,000	5,063,000
\$/3,785 liters (\$/1000 gal.)	16.16	14.17	13.87
Land requirement			
m^2	1,210	5,410	9,060
(ft^2)	(13,000)	(58,200)	(97,500)

a. See Appendices C and D for basis of estimate. All prices are effective as of December, 1980.

b. Because of a 3.5:1 dilution ratio, actual flow through the system is 3.5 times the feed flow rate listed.

Note: Because of a 3.5:1 dilution ratio, actual flow through the system is 3.5 times the feed flow rated listed.

Sludge will be recycled from the clarifier back to the aeration tank by single pumps. A spare pump would be provided for emergency use in the event of a pump failure or for maintenance shutdowns. Waste sludge would be pumped to sludge disposal using a bleed valve on the return sludge line. There is no provision for a sludge disposal system. Recycle system piping, valves, and controls are included in the costs.

The chemical supply system for this stage would consist of a defoamer pump, a polymer pump, sodium hydroxide pump, PAC slurry pump, and sodium carbonate pump, all with associated tankage and piping. Defoamer would be pumped at a rate equivalent to a dose of 2 mg/l. Caustic soda would be pumped at rates equal to those required for the pilot plant operation, shown in Table 52. A pH meter would control the amount of caustic soda added. Sodium carbonate would be added at a rate of 1,840 mg/l. Powdered activated carbon would be added as a slurry with a 0.12 kg/l (1 lb/gal.) concentration. The rate of PAC addition would be in the proportion of 1:7 times the quantity of sludge wasted per day. Polymer would be pumped at a rate equivalent to a dose of 5 mg/l.

All chemical pumping systems would be furnished with spare pumps to be used in emergencies, or during maintenance shutdown of the primary pumps. Mix tanks for the powdered chemicals, such as polymer, sodium carbonate, and PAC would be large enough to require mixing a batch of slurry only once per day. Mixers would be provided in the slurry stock tanks. Defoamer would be pumped from the shipping barrels. Large storage tanks for sodium hydroxide would be provided to contain bulk shipments. The sodium hydroxide would be pumped from the bulk storage tanks to minimize handling. Space would be provided to store a 30 day supply of dry chemicals.

The electrical control system would consist of a motor control center and a supervisory control console. Costs include these panels plus all conduits, wire, and installation labor.

A building would be required to contain the pumps, blowers, chemicals, controls, and a small laboratory space. The costs include building space for the equipment associated with this stage.

Equipment Costs--Using the design parameters developed in Table 52, Stage II nitrification system costs were computed for each of the design flows. Costs for system components were based on vendor quotes and from data in a recent edition of the Dodge Manual for Building Construction, Pricing, and Scheduling (7). Equipment costs include:

1. Concrete aeration tanks.
2. Aeration piping, diffusers, and blowers.
3. Concrete clarifier tanks.
4. Sludge collection system, piping, and pumps.
5. Chemical feed pumps, tanks, and piping.
6. Motor controls.

The Chilton factor method (8) was used to compute the costs of installation, process piping, instrumentation, facilities, and fees.

Operating Costs--Components of the yearly operating cost include power, chemicals, operating labor, maintenance labor, and maintenance material. Items included in the analysis are discussed as follows:

a. Power costs were assumed to be 3.0¢ per KWH. It was assumed that all system motors, except the spare units, would be in operation 24 hours per day, 365 days per year.

b. The chemical consumption would be as listed in Table 52. Costs of chemicals were based on bulk prices in effect as of December 1, 1980. These prices were:

   Polymer: $1.36/kg ($3.00/pound)
   Defoamer: $1.45/ℓ ($5.50/gal.)
   Caustic Soda: $0.38/ℓ ($1.46/gal.) (50 percent solution)
   PAC: $0.13/kg ($0.28/pound)
   Sodium carbonate: $5.45/100 kg ($12.00/100 pounds)

c. Operation labor would be required to oversee the system continuously, regardless of size. It was assumed that a skilled operator earning $19,000 per year would be present during the second and third shifts. The above wages are based on a 5 day work week. Since the system must be operated 7 days per week, additional labor costs were included for continuous operator presence. Assuming that a carbonaceous removal system will be used with this system, one half of the operator charges were allocated to this system.

d. Maintenance labor includes routine inspection, repairs, and preventive maintenance.

e. Maintenance material allowances are included for replacement parts. This cost is a function of total capital costs.

Capital Amortization--An equipment service life of twenty years, amortized at 7.125 percent per year was assumed.

Items Not Included--Examples of items not included in this analysis are: fringe benefits, supervision, overhead, laboratory support, insurance, and administration expenses.

Cost comparisons are calculated on the basis of equivalent annual values which include both capital amortization and operating costs. Results shown in Table 53 are expressed as total annualized costs and costs per 3,785 liters (1,000 gal.) design flow.

Absolute Cost Consideration--
The cost estimates presented in Table 53 are internally consistent and suitable for comparison purposes within the scope of this report. It should be realized that the costs are limited to items listed previously and do not represent the absolute value of total installed cost. The intent is to compare the various alternatives with detail not available in published cost curves.

Land Requirements--Also tabulated in Table 53 are the space requirements for the three designs. Space is provided for the major structures, such as aeration basins, clarifiers, and buildings plus access space. The cost for facilities includes sidewalks and driveways which would occupy some of the access space around the structures.

Equipment Description, (Stage II, Design No. 2)--The Stage II equipment for Design No. 2 would be of the same type as required for Design No. 1, however, sizes and quantities would decrease. All of the design calculations were recomputed on the basis of lower ammonia concentration in the influent wastewater. Although the influent hydraulic flows are identical, tank sizes, power, and quantity of waste sludge could be reduced, due to the absence of dilution water and lower ammonia loading. Design parameters are shown in Table 54.

TABLE 54. STAGE II DESIGN PARAMETERS

Parameter	Recommended range	Design No. 1	Design No. 2
MLVSS, mg/l	2,000-12,000	7,000	7,000
F/M kg NH_3/day/kg MLVSS	0.02	0.019	0.016
Aeration detention time (HRT), hours	24-35	34.8	34.8
Sludge age, SRT, days	7-30	27	27
Sludge production, pounds per day (For 3,785 m^3/day plant)	-	8,294	2,358
Recycle rate, percent	-	51	51
Settling rate, ft/min	-	0.1	0.1
Oxygen required, kg O_2/kg NH_3	3.9	3.9	3.9
Return sludge concentration, mg/l	-	20,000	20,000

Note: ft/min x 0.3048 = m/min

Equipment sizes, quantity, and other descriptive information is shown in Table 55. Equipment costs are given in Table 56. All of the previous discussion related to Stage II, Design No. 1, also applies to Design No. 2.

RESULTS OF THE ACTIVATED CARBON STUDY

Introduction

A granular activated carbon system was used as a post-treatment step to the two stage biological pilot plant. The purpose of the activated carbon study was to investigate conventional and toxic pollutant removal efficiency across the carbon bed. The testing was performed near the end of the on-site field investigations, during a time when the biological system was at steady-state and the second stage bio-oxidation unit was achieving consistent nitrification. Because of the short test duration, carbon breakthrough did not occur.

The testing was conducted between January 31, 1980 and February 7, 1980. Effluent discharging from the second stage activated sludge clarifier was pumped to the carbon column at an average rate of 3.00 liter/min (0.79 gpm). The single carbon column used in the study was operated in the downflow mode and contained 0.14 m^3 (5 ft^3) of regenerated bituminous carbon. Regenerated carbon was used to more closely predict pollutant removal efficiencies in a full-scale system. The resulting contact time was 46.8 minutes. During the study, a total of 15,147 liters (4001.9 gal.) were passed through the activated carbon column. It is realized that the contact time used was liberal; however, flow was limited by upstream processes.

General Results

Table 57 presents the results obtained during the study. Field operating data and conventional pollutant analysis are included. For the three samples collected, influent TOC to the carbon column averaged 125 mg/l while the effluent average concentration was 59 mg/l. The TOC removal efficiency was calculated to be 52.3 percent. The carbon system was also effective in reducing suspended solids in the effluent. The average influent suspended solids concentration was 144 mg/l, while the average effluent concentration resulting from downflow operation of the carbon was 42 mg/l. This is equivalent to a 71 percent removal. However, since the primary function of activated carbon is to remove soluble organics and not suspended solids, any full-scale design should consider upstream filtration and/or carbon column operation in the upflow expanded bed mode. This would reduce the potential for clogging of the bed, resulting from upsets in the biological system; improve pore site availability for organics removal; and reduce backwash requirements.

BOD removal across the carbon column averaged about 41 percent. There were no significant reductions in any of the other BAT parameters analyzed. A separate discussion of the priority pollutant removal data is presented in Section 7 of this report.

TABLE 55. EQUIPMENT DESCRIPTION - STAGE II - DESIGN NO. 2

Flow rate, m^3/day (mgd)	379 (0.1)	1892 (0.5)	3785 (1.0)
Total aeration volume, m^3 (ft^3)	550 (19,400)	2740 (96,900)	5500 (194,000)
Aeration tank number	2	2	2
Aeration tank size (WxLxH) m (ft)	7x9x5 (22x29x15)	9x33x5 (30x108x15)	9x66x5 (30x215x15)
Total air required, m^3/min (cfm)	12 (420)	60 (2100)	120 (4200)
Blower number	3	3	3
Defoamer[a], l/day (gpd)	0.8 (0.2)	3.8 (1.0)	7.6 (2.0)
Sodium hydroxide[b], l/hr (gph)	7.6 (2)	38 (10)	76 (20)
Powdered carbon[c], l/hr (gph)	5.3 (1.4)	26 (7)	53 (14)
Sodium carbonate[d], l/hr (gph)	230 (60)	1100 (300)	2300 (600)
Polymer[e], l/hr (gph)	87 (23)	416 (110)	870 (230)
Total clarifier area, m^2 (ft^2)	17 (185)	86 (925)	170 (1850)
Clarifier number	1	1	2
Clarifier diameter, m (ft)	4.6 (15)	10.7 (35)	10.7 (35)

a. Based on 2 mg/l dosed.
b. Based on 50% concentration.
c. Based on 143 mg/l dosed and 0.12 kg/l stock.
d. Based on 1840 mg/l dosed and 0.12 kg/l stock.
e. Based on 5 mg/l dosed and a stock solution of 1 mg/ml.

TABLE 56. COST SUMMARY - STAGE II - DESIGN NO. 2

Design flow, m^3/day (mgd)	379 (0.1)	1892 (0.5)	3785 (1.0)
Capital costs, $[a]			
Wastewater pumps	7,000	7,000	8,000
Chemical pumps	12,000	17,000	19,000
Chemical mixers	3,000	4,000	7,000
Chemical tanks	8,000	21,000	34,000
Aeration tanks	24,000	92,000	168,000
Aeration piping	4,000	25,000	65,000
Aeration blowers	10,000	32,000	48,000
Clarifier	35,000	63,000	126,000
pH controller	4,000	4,000	4,000
Electrical panel	7,000	8,000	11,000
Delivered equipment total	114,000	273,000	490,000
Equipment installation	46,000	109,000	196,000
Installed equipment total	160,000	382,000	686,000
Process piping	48,000	115,000	206,000
Instrumentation	8,000	19,000	34,000
Capital equipment total	216,000	516,000	926,000
Facilities & fees	302,000	722,000	1,296,000
Capital cost total	518,000	1,238,000	2,222,000
Operating cost, $			
Electrical	7,000	29,000	54,000
Operator	36,000	36,000	36,000
Maintenance labor & materials	11,000	26,000	46,000
Chemicals	108,000	542,000	1,077,000
Yearly operating subtotal	162,000	633,000	1,213,000
Capital amortization			
20 yr life @ 7.125%	49,000	118,000	212,000
Total Annualized cost, $/yr[b]	211,000	751,000	1,425,000
$/3,785 liters ($/1000 gal.)	5.78	4.12	3.90
Land requirement			
m^2	160	1,550	2,750
(ft^2)	(1,760)	(16,700)	(29,600)

a. See Appendices C and D for basis of estimate. All prices are effective as of December, 1980.

b. Annualized cost = operating cost + capital amortization.

TABLE 57. ACTIVATED CARBON STUDY BIOLOGICAL RESEARCH SITE

Date	Cum. Time, Hrs:Min	Flow, gpm	Cum. Flow, gal.	Column Press (psi) Inf.	Column Press (psi) Eff.
1/31/80	0	0.75	-	-	-
	3:15	0.75	146.3	-	-
	9:05	0.75	408.8	-	-
2/4/80	9:05	0.80	408.8	22	20
	15:35	0.89	720.8	26	22
	21:10	0.72	1019.0	26	20.5
2/5/80	33:05	0.79	1533.8	22	16
	37:35	0.79	1747.1	24	19
	40:35	0.75	1882.1	24	19
	42:25	0.81	1964.6	23	18
	46:35	0.81	2167.1	26	21
2/6/80	57:10	0.83	2681.5	26	21
	60:45	0.83	2860.0	26	21
	60:50	-	2864.2	26	21
	60:50	0.83	2864.2	30	26
	63:40	0.83	3005.3	23	19
	66:00	0.77	3121.5	22	18
2/7/80	78:50	0.77	3714.4	22	15
	81:20	0.80	3829.9	23	16
	84:20	0.80	3973.9	24	17
	84:55	-	4001.9	24	17

(continued)

NOTE: gpm x 3.785 = liters/min
gal. x 3.785 = liters
psi x 704 = kg/m^2

TABLE 57. (continued)

	2/5/80		2/6/80		2/7/80	
	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.
TOC, mg/l	158	59	104	51	114	57
SS, mg/l	340	39	240	58	150	30
VSS, mg/l	220	40	230	50	95	18
Color, mg/l	300/400	-	400/500	120/140	500/600	60/80
TKN, mg/l	31	31	17	<1	17	<1
NH_3, mg/l	11	11	1	<1	2	<1
TP, mg/l	2.6	1.8	4.2	2.9	4.0	3.0
CN_T, mg/l	8.6	3.45	8.0	2.45	7.0	3.15
CN_A, mg/l	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05
Phenol, mg/l	0.4	0.2	0.7	0.6	0.9	0.8
O&G, mg/l	21	20	12	17	14	9
BOD, mg/l	87	6	56	40	24	27
SCN, mg/l	<1	<1	<1	<1	<1	<1

Backwashing was required once during the study. The procedure used was to stop the forward feed flow and backwash using service water, which was applied at a rate of 611 l/min/m^2 (15 gpm/ft^2). Backwashing was continued until the water ran clear (about 30 minutes). The carbon column was then put back into service. A sample of the backwash water was collected after 3 minutes of backwashing. Results are shown as follows.

Backwash Sample (Taken After 3 Minutes)
Suspended Solids = 56 mg/l
Volatile Suspended Solids = 16 mg/l

Initially, the backwash water was black and viscous, but cleared quickly with time. Analysis of the single backwash sample indicated that about 30 percent of the suspended solids contained volatile material.

RESULTS OF THE DUAL MEDIA FILTRATION STUDY

Introduction

Dual media filtration tests were performed on the coke plant bio-oxidation effluent as a polishing operation from February 2 through February 11, 1980. The filtration media used in the tests consisted of 0.94 mm effective size anthracite coal placed over 0.47 mm effective size red flint filter sand. A filter media sieve analysis has been shown previously in Figure 14.

Filtration of Bio-oxidation Effluent

The results of filtering bio-oxidation effluent with the small pilot size dual media filter are shown in Table 58. Suspended solids removal rates varied from 17 - 64 percent, averaging 37 percent removal. There was no appreciative removal of oil and grease. Volatile suspended solids were 77 percent of the average influent suspended solids, and 65 percent of the average effluent suspended soilds. These data appear to indicate that biomass floc composed a major portion of the dual media filter effluent. The absence of any oil and grease removal indicates that the oil and grease is free and not associated with the suspended solids.

Data Collection

In addition to overall performance data, the effect of specific operating parameters was determined for each filter run. Since suspended solids removal is the most important measure of filter performance, this parameter was monitored during each run in addition to the filtration rate, headloss and run time.

TABLE 58. DUAL MEDIA FILTRATION OF BIO-OXIDATION EFFLUENT

	Suspended solids, mg/l	Volatile suspended solids, mg/l	Oils and grease, mg/l
		INFLUENT	
Average	190	146	8
Maximum	475	295	12
Minimum	48	47	2
		EFFLUENT	
Average	128	83	9
Maximum	395	230	14
Minimum	32	25	4

Accumulation of solids causes increasing pressure drop across the media. When sufficient solids have been deposited in the media, the headloss exceeds the design influent head and the filter must be backwashed to maintain the desired filtration rate. During any pilot study, the feed suspended solids varies from run to run and rarely equals the design conditions. Therefore, the most meaningful parameter of filter performance is the amount of solids trapped by the filter per unit area per unit headloss. This is usually reported in terms of kilograms suspended solids/m^2/m H.L. (pounds suspended solids/sq ft/ft H.L.).

Appendix B-5.1 presents the suspended solids analysis of all influent and effluent composite samples during each filter run. Also shown in this table is information on filtration rate, headloss, run time, volatile suspended solids concentration, and oil and grease concentration. These data have been summarized in Appendix B-5.2. In this table, calculations have been performed to determine suspended solids removal and solids loading for each of the runs.

Discussion of Results

The data tabulated in Appendices B-5.1 and B-5.2 were used to characterize the filter operation for design purposes. From the data analysis, it has been determined that the average influent suspended solids of 190 mg/l was representative of this influent. Individual filter run influent suspended solids are plotted against effluent suspended solids in Figure 41. It can be seen from Figure 41 that the average suspended solids removal efficiency is 37 percent.

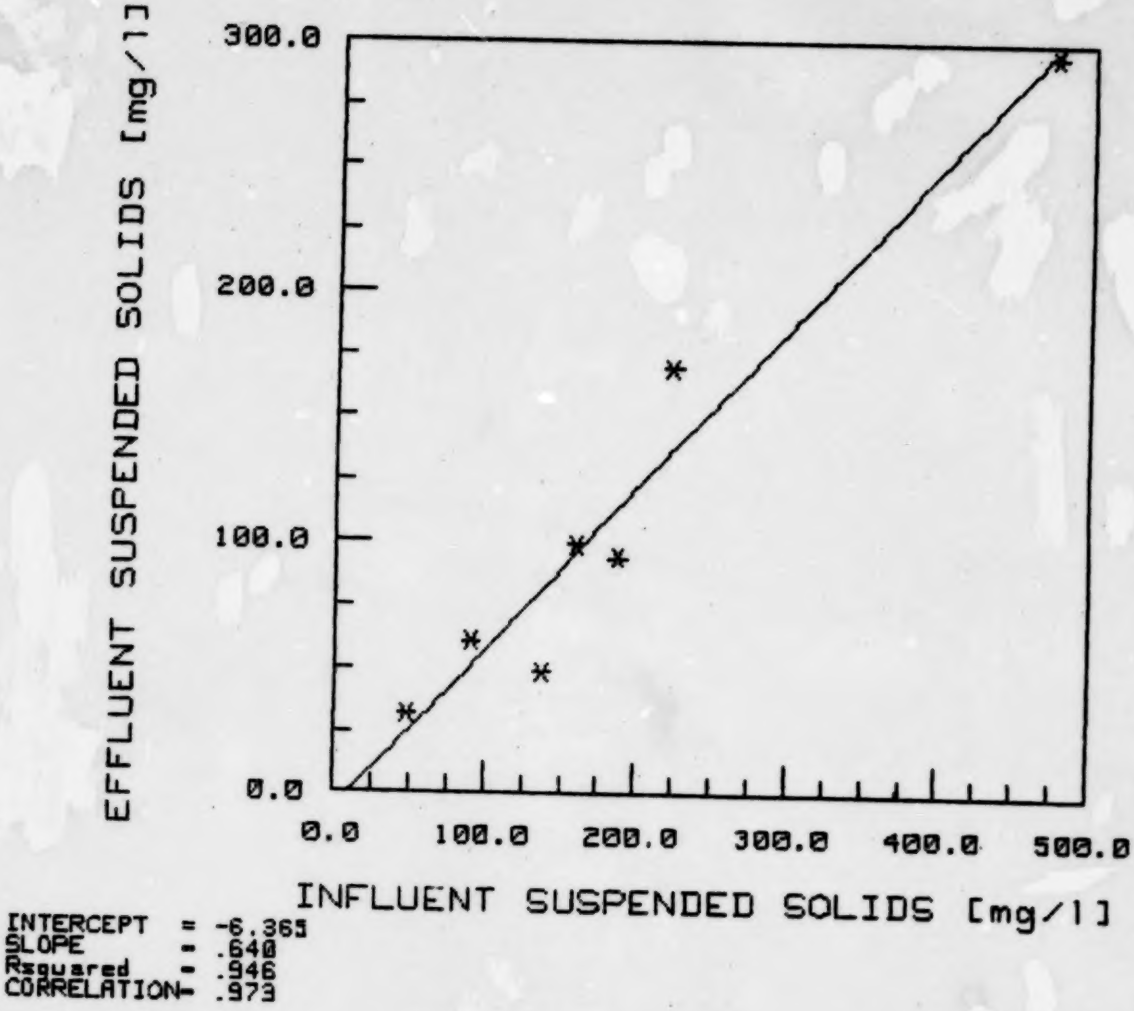


Figure 41. Dual media filter relationship of effluent to influent suspended solids with coke plant bio-oxidation effluent.

Appendix B-5.2 summarizes the average loading for various filtration rates. As seen in the table, solids loadings varied at different flow rates. The solids loading capacities are low, indicating that the filter run lengths will be short. The table indicates that the maximum expected solids loading capacity of the filter is 3.2 kg SS/m^2/m H.L. (0.20 lb SS/ft^2/ft H.L.) while filtering this wastewater.

Basis for Optimum Filtration Rate Selection

Filtration Rate--

Assuming an effluent suspended solids limit of 86 mg/l, the removal efficiency required for the average influent suspended solids level of 190 mg/l SS would be 55 percent. By reviewing the removal efficiency (percent) determined in Appendix B-5.2, it can be seen that the maximum suspended solids removal (50-64%) occurs at filtration rates between 203.7-325.9 l/min/m^2 (5-8 gpm/ft^2). From the pilot data obtained it is obvious that a removal efficiency of >55 percent would be difficult to consistently achieve in a full-scale system. Direct filtration with polymer addition would most likely enhance suspended solids removal and would be recommended for a full-scale installation. Because of the short test duration, it was not possible to investigate the use of polymer during the pilot filtration study. For full-scale design, a loading rate of 244.4 l/min/m (6.0 gpm/ft^2) and a suspended solids removal efficiency of 55% will be used.

Backwash Rate--

Throughout the filtration program, the filter was backwashed after each run until the backwash discharge was clear. Backwash rates varied from 611-1,140 l/min/m^2 (15-28 gpm/ft^2), and backwash times varied between 8-15 minutes. Samples of backwash discharge were collected during one backwash cycle and analyzed for suspended solids. These data are plotted in Figure 42. As seen from Figure 42, nearly all of the suspended solids can be flushed from the media by backwashing for at least 8 minutes. The higher backwash rates produced clear backwash discharge in shorter backwash periods. The total water volume used per backwash ranged from 63 l/m^2 (180 gal/ft^2) to 101 l/m^2 (288 gal/ft^2). The average backwash volume was 74 l/m^2 (212 gal/ft^2).

For design purposes, a conservative backwash rate of 855 l/min/m^2 (21 gpm/ft^2) should be applied for 10 minutes to assure good filter clean out.

Filter Run Duration

The length of time a filter operates before requiring backwashing must be known for design purposes. Due to the time constraints of this study, the maximum filter operating period was 9 hours before backwashing.

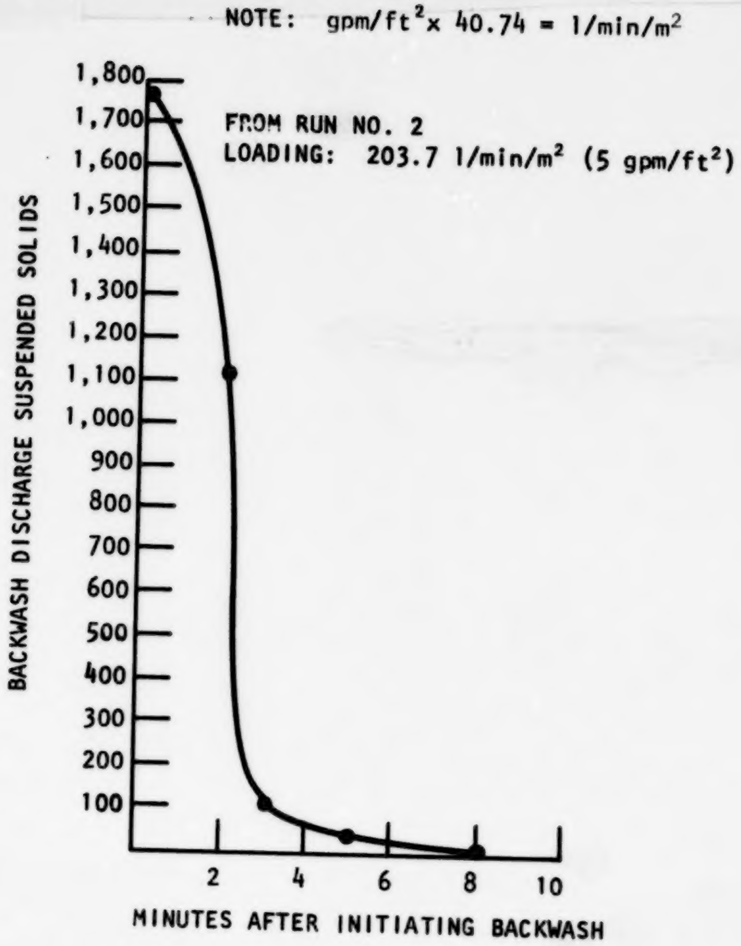


Figure 42. Typical suspended solids washout from dual media filter during backwash.

During the study, at 203.5 l/min/m^2 (5.0 gpm/ft^2) a headloss of 4.21 m (13.8 ft) resulted after 7.5 hours. Assuming a terminal headloss of 9.14 m (30.0 ft) would project a probable run length of 16.3 hours. For a design flow rate of 244.4 l/min/m^2 (6.0 gpm/ft^2) the anticipated run time would be slightly shorter because of a higher total accumulated suspended solids. For design purposes a concervative run length of 12 hours will be used. In filtration practice, this is considered the minimum acceptable length of time between backwashes to prevent excessive usage of backwash water.

Consistent achievement of suspended solids removal may require the addition of polymer upstream. However, the chemical addition system has not been included in the full scale cost estimates. A typical plot of headloss against media bed depth is shown in Figure 43. The curve is taken from run no. 4 after 5.5 hours of operation.

The curve indicates that the majority of the headloss occurs in the sand layer. The flatness of the curve in the top portion of the sand layer indicates that only partial in-depth filtration is occurring at this point in the run.

Metals Removal

During one of the dual media filtration runs, samples of influent, effluent, and backwash were collected for metals analysis. The metals analysis data are summarized in Table 59.

As seen in Table 59, removal of copper (Cu), cadmium (Cd), lead (Pb), and selenium (Se) was low but is more evident by inspection of the backwash data. The difference in influent and backwash concentrations does indicate that significant amounts of these metals were trapped by dual media filtration.

Dual media filtration was effective in removing significant concentrations of zinc (Zn) from the bio-oxidation effluent. The data show that 64 percent of the zinc was removed by filtration. High concentrations of zinc in the backwash also indicates significant removal of this metal.

Concentrations of arsenic (As) and antimony (Sb) were very low in the influent, effluent, and backwash. The data collected showed no measurable removal of arsenic or antimony.

Full-scale Descriptions, Cost, and Space Requirements

The data generated in the pilot scale study were analyzed for system scale-up to full-size installations.

Design Parameters--

From the above pilot study information, parameters were selected for pressure filtration design. Table 60 summarizes the design basis used for the cost estimates. Pressure filtration is defined as having a terminal headloss of 9.1 meters (30 ft). Pressure filter design will be based on constant head-constant rate.

Equipment Description - Pressure filtration--

The pressure filter system for the 379 m^3/day (0.1 mgd) flow would consist of one vertical carbon steel 34 kg (75 lb) code vessel. The system would be provided with an upstream reservoir to accumulate incoming influent during the backwash cycle. The system would also be provided with an effluent tank that would serve as a source of backwash water.

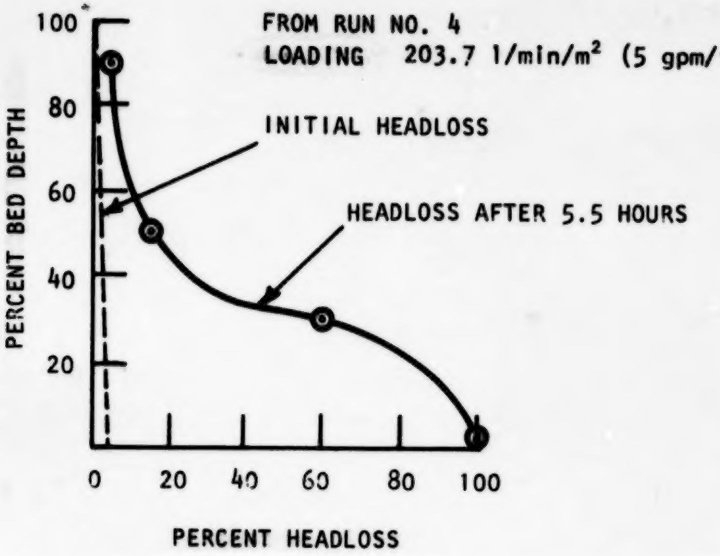


Figure 43. Dual media filter headloss at various media bed depths.

TABLE 59. DUAL MEDIA FILTRATION HEAVY METALS REMOVAL

Sample point	Heavy metal concentration (as total metals) mg/l						
	Cu	Cd	Pb	Zn	As	Se	Sb
Influent	0.06	0.03	<0.1	2.4	<0.0005	0.56	<0.05
Effluent	0.05	0.03	<0.1	0.9	<0.0005	0.42	<0.05
Backwash	0.86	0.13	0.20	30.0	<0.0005	2.33	<0.05

The pressure filter systems for the 1,892 m^3/day (0.5 mgd) and 3,785 m^3/day (1.0 mgd) flow would consist of three vertical carbon steel 34 kg (75 lb) code vessels. During backwash the two pressure vessels in service would serve as the source of backwash water for the out of service vessel.

All systems would be capable of being backwashed first with air at 1.5 sm^3/min/m^2 (5 scfm/ft^2) and then with water at 855 l/min/m^2 (21.0 gpm/ft^2).

TABLE 60. DUAL MEDIA PRESSURE FILTER DESIGN DATA

Type of waste	System design criteria		
Design flow, m^3/day (mgd)	379 (0.1)	1,892 (0.5)	3,785 (1.0)
Filter type	Pressure	Pressure	Pressure
Design influent, mg/l S.S.	190	190	190
Design effluent, mg/l S.S.	86	86	86
Filtration ratio, l/min/m^2 (gpm/ft^2)	244.4 (6.0)	244.4 (6.0)	244.4 (6.0)
Area required, m^2 (ft^2)	1.19 (12.8)	5.94 (63.9)	11.89 (128)
Filter diameter, m (ft)	1.22 (4.0)	1.60 (5.25)	2.29 (7.5)
Filter height, m (ft)	2.44 (8.0)	2.44 (8.0)	2.44 (8.0)
Filter configuration	Vertical	Vertical	Vertical
Number of filters	1	3	3
Area per filter, m^2 (ft^2)	1.16 (12.6)	2.01 (21.6)	4.10 (44.2)
Backwash usage, %	8.6	8.6	8.6
Backwash cycle duration, min	30	30	30
Filter cycle time, hr	11.5	11.5	11.5
Depth of anthracite, cm (in.)	50.8 (20)	50.8 (20)	50.8 (20)
Depth of silica sand, cm (in.)	43.2 (17)	43.2 (17)	43.2 (17)
Depth of gravel, cm (in.)	30.5 (12)	30.5 (12)	30.5 (12)

All systems would have automatically operated butterfly valves, stainless steel diffusers, carbon steel piping, air blower, and one filter feed pump per vessel with one spare pump.

The filters would be filled with 30 cm (12 in.) of graded silica gravel, 43.2 cm (17 in.) of silica sand, and 50.8 cm (20 in.) of anthracite.

A front view cross section of a pressure filter is shown in Figure 44.

Equipment Costs--
Costs for filter components were based on vendor quotations, and from a recent edition of the Dodge Manual for Building Construction, Pricing and Scheduling (7). A cost summary is presented in Table 61. Equipment costs include:

1. Filter vessels and components
2. Backwash reservoir where needed
3. Valves, actuators, and controls for automatic operation
4. Media
5. Pumps and motor control center
6. Air blower for air scour
7. Control console
8. Interconnecting piping between filter vessels where needed
9. The increased cost for the treatment of the backwash water has been included in the clarifier section.

Annualized Costs--
Components of the annualized cost include power, operation labor, maintenance labor, maintenance material, and amortization of capital. Items included in the analysis are discussed as follows:

Electrical: For all filters, the costs of operating the pumps and air scour blowers were based on the backwash frequency and the cycle duration. Electrical costs for pressure filtration include power costs for the feed pumps operating at the design flow, 24 hours per day. Power costs were assumed to be 3.0¢/KWH.

Operation Labor--
Operation of the 379 m^3/day (0.1 mgd) and 1,892 m^3/day (0.5 mgd) system was assumed to require 40% attention during the three shifts while operation of the 3,785 m^3/day (1.0 mgd) system was assumed to require 50% attention during the three shifts. The first shift operator was assumed to be skilled labor grade at $19,000 per year, while second and third shift operators would be unskilled at $16,000 per year.

Maintenance Labor--
This item includes taking all filters out of service once per year for cleaning and inspection. Other costs would include routine replacement of components and general preventive maintenance.

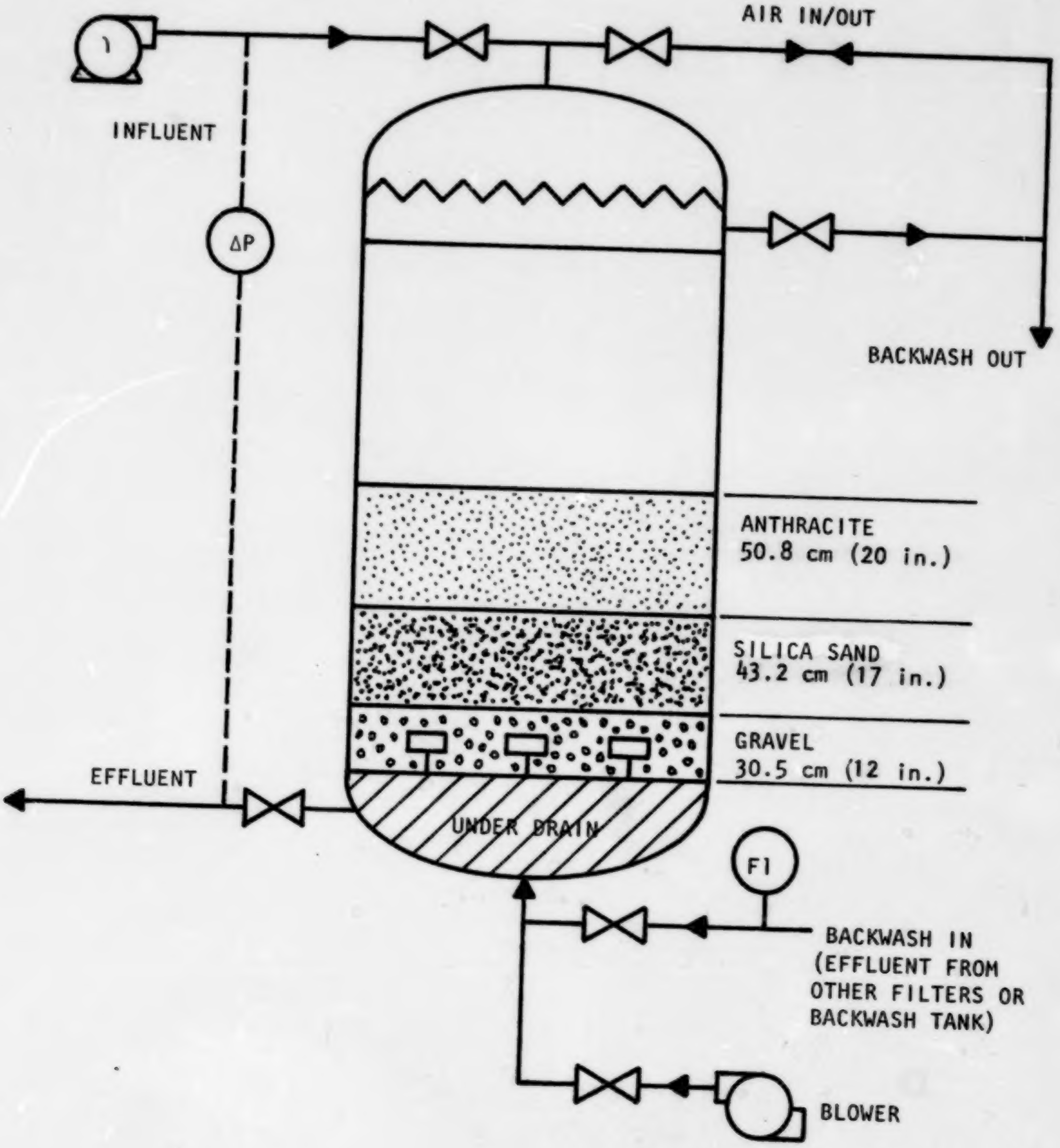

AIR IN/OUT
INFLUENT
ΔP
BACKWASH OUT
ANTHRACITE
50.8 cm (20 in.)
SILICA SAND
43.2 cm (17 in.)
GRAVEL
30.5 cm (12 in.)
EFFLUENT
UNDER DRAIN
FI
BACKWASH IN
(EFFLUENT FROM
OTHER FILTERS OR
BACKWASH TANK)
BLOWER

Figure 44. Front view of vertical pressure filter.

Maintenance Materials--
Included are allowances for replacement parts and replacement of lost media.

Capital Amortization--
An equipment service life of twenty years, amortized at 7.125% per year was assumed.

Items Not Included--
Examples of items not included in this analysis are - fringe benefits, supervision, overhead laboratory support, insurance, and administration expenses. It is reasonable to expect such factors could double the total operating costs.

Labor costs are somewhat independent of filter size and flow loading and can be considered fixed operating costs. Material for maintenance is a function of total capital costs.

Cost comparisons are calculated on the basis of equivalent annual values which include both capital amortization and operating costs. Results shown in Table 61 are expressed as total annual costs in dollars per 3,785 liters (one thousand gallons) design flow.

Absolute Cost Considerations--
The cost estimates presented in Table 61 are internally consistent and suitable for comparison purposes within the scope of this report. It should be realized that the costs are limited to items listed previously and do not represent the absolute value of total installed cost. The intent is to compare the various alternatives with detail not available in published cost curves. For this reason, it is recognized that the absolute costs reported may be less than the published costs for granular media filtration.

Land Requirements--
Also tabulated in Table 61 are the space requirements for the three systems. For the 379 m^3/day (0.1 mgd) system, the filter, pumps, and blower were located along one wall and the influent and the effluent tanks along the opposite wall. Adequate spacing is provided between filter, tanks, and walls to permit easy access to interconnecting piping and valving. An aisle-way is provided down the center of the building.

For the 1,892 m^3/day (0.5 mgd) and the 3,785 m^3/day (1.0 mgd) systems, the filters are located along one wall with the pumps and blower between the filters. A wide aisle-way is provided in front of the filters for servicing.

TABLE 61. DUAL MEDIA FILTER COST SUMMARY

Design flow, m^3/day (mgd)	379 (0.1)	1,892 (0.5)	3,785 (1.0)
Capital Cost[a]			
Filter components	$ 12,000	$ 35,000	$ 58,000
Feed pumps	2,000	4,000	7,000
Upstream reservoir	2,000		
Backwash pump & tank	3,000		
Delivered equipment total	$ 19,000	$ 39,000	$ 65,000
Equipment installation	8,000	16,000	26,000
Installed equipment total	$ 27,000	$ 55,000	$ 91,000
Process piping	8,000	17,000	27,000
Instrumentation	1,000	3,000	5,000
Capital equipment total	$ 36,000	$ 75,000	$123,000
Facilities & fees	78,000	162,000	264,000
Total plant cost	$114,000	$237,000	$387,000
Operating Cost			
Electrical	$ 1,000	$ 3,000	$ 5,000
Operators	20,000	20,000	25,000
Maintenance labor & materials	2,000	4,000	6,000
Yearly operating total	$ 23,000	$ 27,000	$ 36,000
Capital Amortization			
(20 yr life; 7.125%)	$ 11,000	$ 23,000	$ 37,000
Total Annualized Cost			
¢/3,785 liters (¢/1,000 gal.)	93.2	27.4	20.0
Land Requirement			
m^2 (acres)	90 (.022)	100 (.025)	140 (0.34)

[a] See Appendices C and D for basis of estimate. All prices are effective as of December, 1980.

[b] Annualized cost = operating cost + capital amortization.

SECTION 7

PRIORITY POLLUTANT DISCUSSION

INTRODUCTION

Priority pollutant analyses were performed on coke plant wastewater treated in a two stage biological system followed by an activated carbon adsorption unit. A schematic of the wastewater treatment system is given in Figure 45.

A total of 13 samples plus appropriate blanks were collected during February, 1980. The priority pollutant sampling was conducted by Davis and Floyd Engineers, Inc., under subcontract to Southern Research Institute. Priority pollutant analyses were performed by Southern Research Institute (SRI) and the traditional analyses done by Davis and Floyd. All sampling and analysis was performed in accordance with the EPA screening protocol. Appendix B-2 presents the priority pollutant analytical results. A copy of the sample crew log sheets detailing the sampling operation are provided at the end of Appendix B-2.

RESULTS

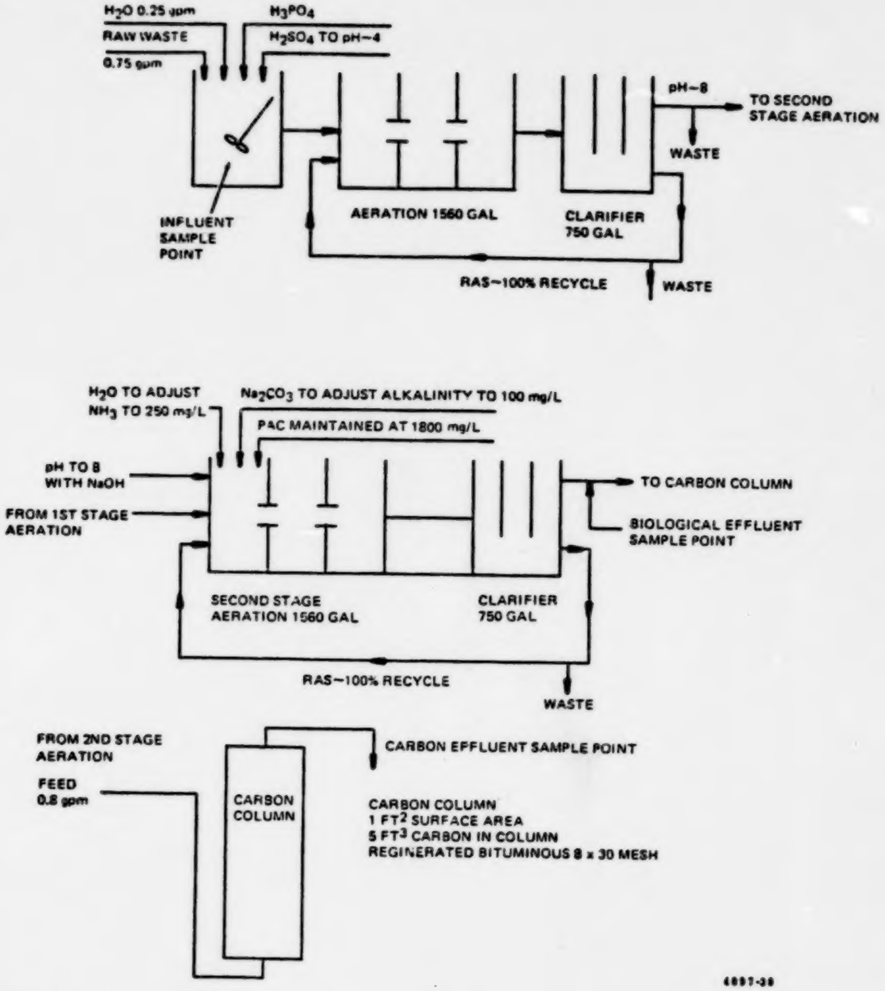
This section describes the results of the priority pollutant analysis performed on the coke plant wastewater treated in a two stage biological system followed by activated carbon adsorption. A site reference sampling code for the samples taken is given in Table 62.

TABLE 62. SAMPLE REFERENCE CODE

Date	Biological influent	Biological effluent	Carbon effluent	Raw water[b]
2-4-80	01678[a]	06179[a]	06180[a]	06181[a]
2-5-80	06182	06183	06184	-
2-6-80	06185	06186	06187	-
2-7-80	06188	06189	06190	-

a. Grab samples.

b. Tap water used for dilution.



H_2O 0.25 gpm
RAW WASTE 0.75 gpm
H_3PO_4
H_2SO_4 TO pH~4
pH~8
TO SECOND STAGE AERATION
WASTE
INFLUENT SAMPLE POINT
AERATION 1560 GAL
CLARIFIER 750 GAL
RAS~100% RECYCLE
WASTE
H_2O TO ADJUST NH_3 TO 250 mg/L
Na_2CO_3 TO ADJUST ALKALINITY TO 100 mg/L
PAC MAINTAINED AT 1800 mg/L
pH TO 8 WITH NaOH
FROM 1ST STAGE AERATION
TO CARBON COLUMN
BIOLOGICAL EFFLUENT SAMPLE POINT
SECOND STAGE AERATION 1560 GAL
CLARIFIER 750 GAL
RAS~100% RECYCLE
WASTE
FROM 2ND STAGE AERATION
FEED 0.8 gpm
CARBON COLUMN
CARBON EFFLUENT SAMPLE POINT
CARBON COLUMN
1 FT^2 SURFACE AREA
5 FT^3 CARBON IN COLUMN
REGINERATED BITUMINOUS 8 x 30 MESH
4097-38

Note: ft x .3048 = m
gpm x 40.74 = lpm

Figure 45. EPA pilot plant schematic
Follansbee, W.V.

Metals

Priority pollutant metal verification data are presented in Table 63. The activated sludge system influent contained significant concentrations of zinc and selenium. The raw water added to the treatment system as makeup water had a high zinc concentration (630 μg/l). The analysis program did not differentiate whether the source of the high zinc influent was the raw waste or due to the makeup water. Influent arsenic and silver concentrations were generally low.

Zinc in the activated sludge system effluent showed a threefold increase while selenium concentrations decreased threefold from influent levels. Arsenic levels were reduced threefold in the activated sludge system. Both arsenic and silver concentrations in the activated sludge effluent were extremely low.

Zinc concentrations in the carbon column effluent were reduced threefold from the activated sludge effluent levels. The carbon column effluent zinc concentration was 300 μg/l to 560 μg/l during the sampling period. SRI reported a zinc quality assurance spike recovery of 92 percent. The selenium concentration in the carbon column effluent decreased approximately 25 percent from activated sludge system effluent levels. Selenium concentrations were 250 μg/l to 420 μg/l in the carbon column effluent. A spike recovery of 115 percent was reported for selenium. Arsenic and silver concentrations in the carbon column effluent were very low. The mean silver concentration was 12 μg/l with a reported spike recovery of 94 percent. Arsenic levels ranged from 82 μg/l to 130 μg/l in the carbon column effluent. SRI reported a 97 percent arsenic spike recovery. Carbon column effluent concentrations of all the other metals were less than the detection limit except for mercury. Mercury concentrations of 0.2 μg/l and 0.6 μg/l (detection level 0.1 μg/l) were reported in the final carbon effluent.

Organics

Organic priority pollutant analytical data are presented in Table 64. Three groups of organic priority pollutant compounds were verified, namely 1) volatile organics, 2) base/neutral extractable organics, and 3) acid extractable organic compounds.

Volatile Organics--

The influent to the activated sludge system contained significant concentrations of the volatile organics toluene and benzene. All the other volatile organics in the influent were at or below the detection limit except for methylene chloride. Reported methylene chloride concentrations in the influent were 489 μg/l to 589 μg/l. Toluene, benzene, and methylene chloride concentrations in the activated sludge system effluent generally less than the detection limit. SKI reported volatile organic spike activated sludge clarified effluent and the carbon column effluent were generally less than the dection limit. SRI reported volatile organic spike recoveries of 74 percent to 104 percent for the activated sludge system influent and 80 percent to 125 percent for the carbon column effluent.

TABLE 63. VERIFICATION REPORT FOR METALS

Metals		Activated sludge system influent			Activated sludge system effluent			Carbon column effluent		
	2/04/80	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80
Ag	<10	18	18	14	12	12	12	12	12	12
As	<10	350	340	390	86	95	140	86	82	130
Be	<10	<10	<10	<10	<10	<10	<10	<10	<10	<10
Cd	<15	<15	<15	<15	<15	<15	<15	<15	<15	<15
Cr	<10	21	24	25	<10	<10	10	<10	<10	<10
Cu	<50	71	71	57	<50	<50	<50	<50	<50	<50
Hg	<0.1	7.8	3.7	5.6	3.8	2.1	1.2	0.2	0.6	-
Ni	<100	<100	<100	<100	<100	<100	<100	<100	<100	<100
Pb	<10	<10	<10	<10	<10	<10	<10	<10	<10	<10
Sb	<10	260	120	120	<20	<20	<20	<20	<20	<20
Se	<10	1400	1400	1320	330	350	550	250	290	420
Tl	<10	<10	<10	<10	<10	<10	<10	<10	<10	<10
Zn	630	480	460	390	1400	1200	920	300	560	330

Notes: 1. All results in μg/l.
2. (<) represents the lower limit of detection for the method and/or instrument.

TABLE 64. GC/MS VERIFICATION REPORT FOR ORGANICS

Compound	Raw water	Activated sludge system influent			Activated sludge system effluent			Carbon column effluent		
	2/04/80	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80
1. Volatile Organic										
Methylene chloride	15	589	548	489	<10	<10	<10	22	<10	<10
Toluene	<10	722	542	556	<10	<10	14	<10	<10	15
Chloroform	ND	<10	<10	<10	<10	<10	25	<10	<10	31
Trichloroethylene	10	ND	ND	ND	ND	ND	ND	ND	ND	ND
Bromodichloromethane	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Dibromochloromethane	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Benzene	15	9,820	6,100	6,210	ND	ND	ND	ND	ND	ND
2. Base/Neutral Extractable										
Naphthalene	ND	964	760	1,160	<10	ND	ND	ND	ND	ND
Acenaphthylene	ND	59	40	57	ND	ND	ND	ND	ND	ND
Fluorene	ND	48	33	34	ND	ND	ND	ND	ND	ND
Anthracene/Phenanthrene	ND	562	454	582	<10	<10	<10	<10	<10	ND
Fluoranthene	ND	333	311	410	<10	<10	<10	ND	ND	ND
Pyrene	ND	212	196	265	<10	<10	<10	ND	ND	ND
Chrysene/ Benzo(a)anthracene	ND	84	86	100	ND	ND	<10	ND	<10	ND
Benzo(b)fluoranthene	ND	16	12	13	ND	ND	<10	ND	ND	ND
Benzo(a)pyrene	ND	<10	ND	ND	ND	ND	<10	ND	ND	ND
3. Acid Extractable										
2-Chlorophenol	ND	985	1,660	1,170	ND	ND	ND	ND	ND	ND
2-Nitrophenol	ND	29	<25	<25	<25	ND	ND	<25	ND	ND
Phenol	ND	65,400	65,200	55,300	ND	ND	<25	ND	ND	ND
2,4-Dimethylphenol	ND	6,810	6,970	7,600	ND	ND	ND	ND	ND	ND

Notes: 1. All results are reported as μg/l unless indicated.
2. ND = Not Detected

Base/Neutral Extractable Organics--

The activated sludge system influent contained five BAT monitored base/neutral extractable compounds in moderate concentrations. Of these five compounds, naphthalene was present in the greatest concentration. The mean naphthalene concentration in the influent was 960 μg/l. Fluoranthene, pyrene, acenaphthylene and fluorene were also present in moderate concentrations in the influent. Quality control spike recoveries for the influent base/neutral extractable organics of interest were 58 percent to 111 percent.

The activated sludge system reduced all base/neutral extractable organics to less than 10 μg/l. Base/neutral extractable organics were generally below detection levels in the final carbon column effluent. SRI reported spike recoveries of 68 percent to 139 percent in the carbon column effluent.

Acid Extractable Organics--

The acid extractable organic group contains the priority pollutant phenols. Five phenolic compounds were analyzed in the activated sludge system influent. The concentration of phenol in the influent was significant. Substantial concentrations of 2-chlorophenol and 2,4-dimethylphenol were also verified in the influent. SRI encountered analytical interference in the influent phenol quality control spike recovery. Reported recoveries of the other phenolic compounds present in the activated sludge system influent were 17 percent to 120 percent.

The concentration of all phenolic compounds in the activated sludge system effluent were less than detection limits. Likewise no phenol was detected in the carbon column effluent. SRI reported spike recoveries of <10 percent to 45 percent for phenols in the carbon column effluent.

SUMMARY

Metals

The pilot treatment system was not able to effectively reduce selenium and zinc concentrations. Selenium was reduced by 77 percent from a mean influent concentration of 1,370 μg/l to a final effluent of 320 μg/l. Zinc levels were only slightly reduced. The mean influent concentration of zinc was 443 μg/l and the effluent 400 μg/l. Zinc was found in high concentration (630 μg/l) in the raw water added to the treatment system as makeup water. Arsenic was effectively removed. Arsenic was reduced from a mean influent concentration of 360 μg/l to an effluent concentration of 99 μg/l. Influent silver concentrations (18 μg/l) as well as effluent levels (12 μg/l) were low in concentration. All the other priority pollutant metals were removed to concentration levels close to or below detection limits.

Organics

Volatile Organics--

The treatment process was effective in removing all volatile organic priority pollutants to low levels. Toluene was reduced from a mean influent concentration of 607 µg/l to a final effluent of less than 10 µg/l (on two of the three sampling dates). While the activated sludge system influent contained 6,100 µg/l to 9,800 µg/l of benzene, no benzene was detected in the activated sludge or the final carbon column effluents.

All other volatile organic compounds in the activated sludge and carbon column effluents were generally below the detection limits.

Base/Neutral Extractable Organics--

The treatment processes were effective in removing all monitored base/neutral extractable organics. The concentration of all nine of the base/neutral compounds found in the influent were reduced through the treatment process to less than 10 µg/l in the activated sludge final effluent.

Acid Extractable Organics--

All acid extractable organic compounds monitored were effectively removed by the treatment system. The system influent contained a mean concentration of 62 mg/l phenol. Phenol was reduced by the treatment process to less than the detection limit in the final effluent. All other acid extractable organic compounds in the final effluent were below the detection limit.

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APPENDIX A. COKE PLANT PRODUCTION FIGURE AND INTAKE WATER QUALITY

APPENDIX A-1. COKE PLANT PRODUCTION FIGURE.

Date	Coke production (tons/day)	Wastewater discharge (gal/day)	Flow (gal/ton)
10/19/79	4956.15		
20	5532.95	526,032	95.1
21	5364.20	530,352	98.9
22	4614.75	524,160	113.6
23	4450.10	528,480	118.8
24	4325.70	550,512	127.3
25	5188.55	521,424	100.5
26	4809.95	518,544	107.8
27	5219.85	518,688	99.4
28	5396.40	530,064	98.2
29	5525.55	513,216	92.9
30	4624.40	470,736	101.8
31	4819.75	501,840	104.1
11/01	5054.80	476,064	94.2
02	5540.55	462,672	83.5
03	5240.30	505,584	96.5
04	5453.30	492,768	90.4
05	4644.00	498,960	107.4
06	5418.20	524,592	96.8
07	4637.10	516,096	111.3
08	4892.50	516,096	105.5
09	4421.10	518,688	117.3
10	4885.30	518,688	106.2
11	5317.90	523,152	98.4
12	5623.05	519,264	92.3
13	4977.90	502,416	100.9
14	5398.60	502,416	93.1
15	4866.50	501,840	103.1
16	5080.80	523,152	103.0
17	5028.20	537,552	106.9
18	5339.80	570,240	106.8
19	4979.25	589,248	118.3
20	4907.10	437,472	89.2
21	5149.20	502,848	97.7
22	4835.40	502,848	104.0
23	5627.25	500,256	88.9
24	5338.35	502,272	94.1
25	5076.20	504,000	99.3
26	4783.80	487,296	101.9
27	4898.50	537,984	109.8
28	5253.50	495,216	94.3
29	4741.30	495,216	104.4
30	4843.20	432,576	89.3
12/01	5327.05	432,576	81.2
02	5344.75	476,496	89.2
03	5143.30	476,496	92.6
04	4715.70	476,496	101.0
05	3992.70	441,360	110.5
06	4628.50	421,920	91.2
07	5132.30	421,920	82.2
08	4708.70	432,432	91.8
09	4857.05	436,464	89.9
10	4178.40	438,768	105.0
11	5164.70	438,768	85.0
12	4992.60	475,488	95.2
13	5055.45	477,216	94.4
14	5112.15	455,616	89.1
12/15	5621.10	480,528	85.5
16	4965.40	512,784	103.3
17	5274.00	419,760	79.6
18	5121.80	513,072	100.2
19	4393.85	498,528	113.5
20	4579.25	523,872	114.4
21	4519.00	529,056	117.1
22	5022.80	529,056	105.3
23	5181.90	508,752	98.2
24	5446.30	529,488	97.2
25	4927.80	552,240	112.1
26	5134.90	555,120	108.1
27	5044.50	561,168	111.2
28	4793.95	558,000	116.4
29	5188.30	559,008	107.7
30	5005.10	559,008	111.7
31	5347.95	558,864	104.5
1/01/80	4886.80	528,192	108.1
02	4874.85	528,192	108.3
03	4833.75	537,120	111.1
04	5414.30	537,120	99.2
05	5166.40	533,376	103.2
06	5039.00	508,176	100.8
07	4853.15	512,496	105.6
08	5218.90	514,368	98.6
09	5141.40	507,600	98.7
10	4463.75	511,056	114.5
11	4548.40	516,240	113.5
12	5328.95	513,792	96.4
13	5067.80	513,648	101.4
14	5023.55	521,280	103.8
15	4988.50	561,744	112.6
16	4670.95	533,376	114.2
17	4267.95	499,536	117.0
18	4511.75	534,528	118.5
19	4427.45	534,528	120.7
20	5612.10	531,360	94.7
21	4741.80	527,328	111.2
22	4840.65	517,392	106.9
23	4817.15	502,704	104.4
24	4726.15	432,864	91.6
25	4187.35	499,680	119.3
26	5140.60	492,048	95.7
27	5002.05	511,200	102.2
28	4695.00	476,208	101.4
29	4678.95	513,648	109.8
30	4639.30	549,792	118.5
31	4427.15	532,080	120.2
2/01	4213.75	523,584	124.3
02	4881.85	522,144	107.0
03	4606.40	525,312	114.0
04	4557.25	528,768	116.0
05	4878.75	522,864	107.2
06	5103.50	540,720	106.0
07	5250.20	566,928	108.0
08	4757.60		

Note: tons x 0.9078 = metric tons
gal. x 3.785 = liters

APPENDIX A-2. COKE PLANT INTAKE WATER QUALITY

1977 NUS SURVEY

Date	Temp. ($^{\circ}$C)	pH	Total Suspended Solids (mg/1)	Oil & Grease (mg/1)	Ammonia-N (mg/1)	Cyanide Total (mg/1)	Phenol (mg/1)
7/7/77	29	6.8	5	2	0.10	<0.005	0.002
7/13/77	28	6.7	5	2	<0.10	<0.005	0.004
7/21/77	30	6.3	85	4	0.11	<0.005	0.005
7/27/77	20	5.8	106	1	0.10	0.005	<0.001
8/3/77	24	6.0	84	10	0.22	<0.005	0.004
8/9/77	26	6.9	32	1	0.10	<0.005	0.004
8/17/77	-	7.1	20	3	0.10	0.02	0.001
8/23/77	21	6.0	47	2	0.10	<0.005	0.002
9/13/77	22	7.0	7	2	-	-	-
9/21/77	19	7.1	42	<1	-	-	-
9/27/77	23	6.8	38	3	-	-	-
10/7/77	18	7.2	50	2	-	-	-
10/10/77	16	6.7	-	-	-	-	-
10/19/77	12	6.8	<1	1	-	-	-
11/2/77	19	6.8	46	2	-	-	-
11/9/77	16	7.1	7	5	-	-	-
11/14/77	16	6.9	20	3	-	-	-
11/28/77	11	7.1	-	-	-	-	-
12/7/77	5	6.9	-	2	-	-	-
12/12/77	5	6.7	-	2	-	-	-
12/21/77	12	6.4	-	6	-	-	-
12/27/77	3	7.0	-	3	-	-	-

(1) Samples of intake water are collected as composites composed of 4 equal aliquots collected over 24 hours. Temperature and pH are grabs. Samples are collected on a frequency of 4/30 days.

APPENDIX B. COKE PLANT RAW DATA

APPENDIX B-1. CONVENTIONAL POLLUTANT DATA SUMMARY

Date	pH	SS mg/l	VSS mg/l	$S^=$ mg/l	TDS mg/l	Phenol mg/l	O&G mg/l	TKN as N mg/l	NH_3 as N mg/l	CN_T mg/l	CN_A mg/l	TOC mg/l	Total P as P mg/l	BOD mg/l	Color APHA Units	SCN mg/l
System Feed (Diluted Raw)																
10/23/79		30				600		1,030	900	4.6		940	20.1			300
24	8.50	48	31			765		1,120	994	7.6		980	9.6		1,400	
25	8.70	52		42	8,830	880			1,220	4.0		1,660	<0.02			480
26						550			930	6.7	0.4	960	3.6		1,500/2,000	
30						810			2,090			1,000	8.5			350
31						805			1,920			680	12.8			330
11/01/79					6,270	820			1,670	3.4		830	14.5			330
02					4,770	800			1,560	9.9		1,010	19.3			310
03	7.60	45										1,260	16.4			
04	7.90											1,050	9.7			
05	6.90				8,530	800	2		870	7.9		1,140	5.2			370
06	7.50				9,170				810			1,180				
07	8.30	48	32		6,930	695	6		750	12.4		930				360
08	7.35	86	53						1,100			980				
09	7.10	86	56									1,040		1,820		
10	7.40	31	21						1,190			1,030				
11		135	50									1,000				
12	8.35	59	35	32		890		1,590	1,420	10.8	1.0	1,110				385
13												1,180				
14	7.50	110	42					2,010	1,760	10.0	<0.05	1,050		2,240	1,800	410
15												1,050				
16		41	30	34		920	4	1,420	1,370	9.6	0.2	1,250	7.3	2,100		360
17												1,410				
18												1,000				
19		37	32	32	5,520	780	5	793	692	6.5	0.7	1,040	6.2	1,920	600/700	320
20		49	37		5,710							1,040			400/600	
21	7.50	47	33		8,070	850			990	8.7	0.9	1,090		2,110	500	310
22									815			1,070				360
23		48	34		8,650	860		1,512	910	14.0	0.7	1,060			600	
24												1,020				
25		45	30									310				
26	7.70	45		46	10,000	515	3	1,320	1,160	10.3	1.0	1,040		1,900	400	350
27												980				
28	7.30	85	51		6,390	605		1,160	846	9.0	1.0	180		2,280	600/700	375
29												1,300				
30	7.50	58	43	35	7,990		27	1,170	274	10.5	1.9	750		2,070	600/700	420
12/01/79												1,090				
03		50	37			720	18	1,020	851	11.0	1.6	1,140	7.1	2,040		455
04												1,000				
05		53	41			800	45	1,070	944	5.5	<0.05	720	6.3	2,390		440

continued

APPENDIX B-1. (Continued)

Date	pH	SS mg/l	VSS mg/l	$S^=$ mg/l	TDS mg/l	Phenol mg/l	O&G mg/l	TKN as N mg/l	NH_3 as N mg/l	CN_T mg/l	CN_A mg/l	TOC mg/l	Total P as P mg/l	BOD mg/l	Color APHA Units	SCN mg/l
System Feed -(Diluted Raw) continued																
12/06/80												1,180				
07		75	47			870	26	1,110	1,020	10.5	<0.05	850		2,550		
08												565				
09												944				
10		62	37			640	27	722	630	7.2	3.5	1,120	7.4	2,080		420
11												1,480				
12		88	41					700	608	8.9	<0.05	1,500	7.1			
13												1,580			400/600	440
14		95	38	40		800	8	788	650	9.5	<0.05	1,470	6.9	2,030	600/800	385
15												1,430				
16												1,300				
17		66	29			550	6	1,140	526	5.8	<0.05	1,380	12.9	1,490	300/400	390
18												1,180				
19		51	24					661	616	5.6	<0.05	1,400	11.4	1,630	400/500	495
20												1,230				
21		64	36	24		610	16	501	420	4.3	<0.05	1,290	11.1	1,680	300/400	435
22												1,200				
23												1,370				
24	6.90	70	50			580	105	482	319	4.9	<0.05	1,390		1,900	400/500	525
25												1,250				
26	6.80	58	30			510		372	280	0.8	<0.05	1,380	6.3	1,800		545
27												1,300				
28	7.00	64	46	15		560	18	398	336	1.2	<0.05	1,360		1,870	200/300	490
29												1,430				
30												1,410				
31		64	38			530		384	356	4.7	<0.05	1,330				515
1/01/80												1,320				
02	6.90	39	27	28	8,780	640	6	846	745	3.3	1.4	1,500	7.9	2,100	400/600	570
03												1,260				
04		122	59					921	909	12.6	<0.05	1,110	9.3	1,770		600
05												1,140				
06												1,410				
07		97	52			590		694	599	26.9		1,180	9.0	1,812		520
08												1,200				
09		76	31						553	6.5	2.3	1,250	10.5	1,968		610
10												1,260				
11		80	36	25		650	15	470	302	11.6	<0.05	1,200	12.2	1,722	200/300	535
12												1,250				
13												1,140				
14		70	28			440	14	602	510	9.0	12.08	1,100	13.2	1,530		590

continued

Appendix B-1 (continued)

Date	pH	SS mg/l	VSS mg/l	$S^=$ mg/l	TDS mg/l	Phenol mg/l	O&G mg/l	TKN mg/l	NH_3 as N mg/l	CN_T mg/l	CN_A mg/l	TOC mg/l	Total P as P mg/l	BOD mg/l	Color APHA units	SCN mg/l
System Feed (Diluted Raw) continued)																
1/15/80												1,160				
16		58	25					787	694	9.0	7.35	1,150	12.5	1,614		580
17												1,090				
18		35	27	29		540	15	580	500	4.0	<0.05	1,000	12.3	1,371	150/180	550
19												920				
20												960				
21		96	86			520	16	370	300	25.3	<0.05	970	10.8	1,290		580
22												196				
23		64	37					350	280	14.4	<0.05	940	10.0	1,398		550
24												910				
25		74	46	28		490	8	340	220	15.2	<0.05	950	10.2	1,503	180/210	500
26												1,020				
27												1,040				
28		93	47			540	16	616	504	21.2	<0.05	1,050	9.8	1,614		470
29												1,120				
30		67	40			540		605	540	27.9	0.05	1,240	9.3	1,758		550
31												1,160				
2/01/80		77	30	28		560	7	697	636	32.4	5.8	1,197	7.63	1,560	250/300	470
02		88	29	28		440	7	728	652	20.5	<0.05	1,125	8.56	1,458	200/250	475
03		80	27	28		560	4	389	291	4.6	<0.05	1,116	8.00	1,500	150/200	445
04		205	80	30		580	15	375	297	4.2	<0.05	1,170	8.00	1,590	150/200	490
05		160	70	33		580	24	409	316	6.0	1.7	1,197	8.00	1,443	200/300	490
06		30	30	34		600	15	330	288	6.3	2.4	1,200	10.5	1,659	300/400	455
07		20	19	23		540	18	364	283	8.5	<0.05	1,200	10.63	1,623	200/300	425
Activated Sludge I - Clarifier Effluent																
10/25/79	6.50	315		1.6		1.5			1,130	4.5		280	6.0			370
26						1.2			910	4.8	0.4	240			1,000/1,200	
30						12			1,710			276	2.6			360
31						2.7			1,750			255	2.5			350
11/01/79					6,560	26						295	4.5			360
02						38				2.3		345				330
03	6.90	56										160	15.6			
04	6.70											200	13.2			
05	6.40				8,390	3.1	1		760	4.60		200	10.8			350
06	6.10				8,620				840			260				
07	5.70	89	64		8,150	28	<1		620			190				
08	7.00	197	50						794			110				
09	6.95	116	69									248		265		

continued

Appendix B-1 (continued).

Date	pH	SS mg/l	VSS mg/l	$S^=$ mg/l	TDS mg/l	Phenol mg/l	O&G mg/l	TKN as N mg/l	NH_3 as N mg/l	CN_T mg/l	CN_A mg/l	TOC mg/l	Total P as P mg/l	BOD mg/l	Color APHA Units	SCN mg/l
Activated Sludge I - Clarifier Effluent (continued)																
11/10/79		165	91									234				
11		93	53									276				
12	6.50	77	48	2.4		4.1		1,250	1,090	7.2	0.2	320				370
13												260				
14	7.90	189	156					1,600	1,570	10.0	0.5	440		905	1,500	400
15												330				
16		198	158	3.9		2.9	5	1,600	1,360	8.6	0.3	510	5.3	805		380
17												450				
18												630				
19		86	77	3.5	5,500	2.3	5	1,020	955	5.7	<0.05	305	3.2	340	600	350
20		81	70		5,880							300			600	
21	6.40	109	78		8,730	5.0			875	4.7	0.8	315		270	1,500	335
22									630			260				
23		42	39		8,290	3.7		874	820			290			800	
24												240				
25												215				
26	7.60	57	55	6.6	9,000	2.8	2	1,060	970	10.7	<0.05	260		435	1,125	385
27												240				
28	7.50	118	85		7,820	0.1				11.0	<0.05	740		320	1,200/1,600	374
29												265				
30	7.50	88	73	23	8,670		7	1,160	361	8.5	<0.05	250		245	1,200/1,600	400
12/01/79												250				
02																
03		48	41			7.4	8	994	840	5.7	0.4	138	4.2	300		425
04												249				
05		76	66			70	57	1,056	935	5.2	2.8	444	5.7	630		478
06												480				
07		125	99			90	36	1,120	977	4.2	<0.05	590	6.3	785		480
08												480				
09												276				
10		94	66			16	31	899	790	7.8	<0.05	234	1.5	430		445
11												260				
12		184	108					739	720	0.2	<0.05	240	4.7		800/1,200	446
13												230				
14		121	68	17		5.6	4	682	582	0.2	<0.05	280	1.0	130	600/800	440
15												205				
16												135				
17		100	53			3.8	2	1,484	666	5.1	<0.05	185	1.7	145	800/1,000	425
18												145				
19		98	65					622	605	1.8	1.4	175	5.8	145	800/1,000	440
20												190				
21		135	81	4.1		5.7	4	661	437	0.2	<0.05	155	9.0	165	800/1,000	430

continued

Appendix B-1 (continued).

Date	pH	SS mg/l	VSS mg/l	S$^=$ mg/l	TDS mg/l	Phenol mg/l	O&G mg/l	TKN as N mg/l	NH$_3$ as N mg/l	CN$_T$ mg/l	CN$_A$ mg/l	TOC mg/l	Total P as P mg/l	BOD mg/l	Color APHA Units	SCN mg/l
Activated Sludge I - Clarifier Effluent (continued)																
12/22/79												205				
23												215				
24	7.90	150	100			0.8	5	518	353	0.7	<0.05	235		255	1,000/1,200	470
25												250				
26	8.10	102	52			0.4		431	358	0.3	<0.05	200	10.1	135		515
27												195				
28	8.10	123	69	2.1		2.6	5	431	353	0.2	<0.05	110		165	800/1,000	480
29												185				
30												90				
31		111	56			1.4		420	386	9.6	<0.05	40				3
1/01/80												70				
02	8.40	94	81	7.9	8,400	<0.1	3	652	620	11.0	<0.05	35	9.0	32	800/1,200	4.6
03												188				
04		150	80					868	658	20.9			9.4	41		23
05												170				
06												250				
07		155	100			1.2		986	885	19.3		185	9.2	97		390
08												270				
09		98	58						627	25.0	3.0	260	8.8	48		515
10												270				
11		69	53	0.6		0.4	13	540	482	12.5	7.4	240	8.6	35	1,000/1,200	525
12												230				
13												220				
14		101	48			0.6	9	496	440	17.9	9.2	200	10.2	60		410
15												180				
16		139	67					655	610	6.0	<0.05	150	11.2	105		130
17												120				
18		104	60	2.3		0.4	10	820	700	11.1	<0.05	172	11.3	39	400/600	11
19												171				
20												152				
21		75	42			2.1	5	510	500	16.5	<0.05	110	9.4	13		8
22												160				
23		73	47					420	400	22.0	<0.05	176	8.4	61		4
24												172				
25		87	51	1.9		0.5	9	360	330	20.2	<0.05	135	8.7	34	600/800	2
26												112				
27												127				
28		103	54			0.2	12	479	420	24.0	<0.05	135	5.6	175		3
29												332				
30		207	111			1.2		787	700	17.4	<0.05	260	7.8	90		480
31												296				

continued

Appendix B-1 (Continued)

Date	pH	SS mg/l	VSS mg/l	$S^=$ mg/l	TDS mg/l	Phenol mg/l	O&G mg/l	TKN as N mg/l	NH_3 as N mg/l	CN_T mg/l	CN_A mg/l	TOC mg/l	Total P as P mg/l	BOD mg/l	Color APHA Units	SCN mg/l
Activated Sludge I - Clarifier Effluent (continued)																
2/01/80		206	90	5.2		1.4	6	694	619	15.7	5.7	360	6.6	43	1,200/1,400	525
02		198	100	3.6		0.1	4	724	647	15.2	1.4	140	7.1	66	1,250/1,500	470
03		220	106	6.4		0.8	10	661	596	5.6	<0.05	250	8.1	75	1,000/1,200	345
04		280	105	2.2		0.4	10	585	521	16.9	2.9	182	7.7	81	1,000/1,200	100
05		260	140	1.4		1.2	20	487	437	14.7	<0.05	230	7.1	150	1,000/1,200	71
06		120	100	1.2		1.4	9	386	381	8.6	0.6	248	7.8	190	500/600	235
07		70	65	2.8		0.8	12	381	336	7.4	<0.05	266	8.3	155	1,200/1,400	330
Activated Sludge II - Clarifier Effluent																
11/02/79												151	4.5			
05	8.20				8,540	0.1			900	4.2		108	12.1			<1
06	8.10				8,620				950			122				
07	7.75	55	31		8,000	<0.1			770			82	12.2			
08	7.05	86	48						773			85	7.2			
09	7.75											107		54		
10		63	45									84				
11		52	39									102				
12	7.35	55	38	<0.2		1.7		1,130	1,050	7.0	0.9	95				44
13												146				
14	7.60	151	82					1,390	1,350	6.0	0.9	90		165	1,500	
15												93				
16		108	90	1.9		2.4		1,630	1,550	5.0	<0.05	75	4.6	100		310
17												81				
18												69				
19		93	87	1.7	5,890	1.6	2	1,390	1,270	5.9	<0.05	160	3.0	110	1,000/1,200	310
20		76	63		5,820							168			1,000/1,200	
21	8.10	91	65		8,970	<0.1			1,010	5.0	0.4	116		85	800	27
22												168				
23		44	40		8,890	0.2		952	970	6.9	<0.05	100			800	2
24												109				
25												156				
26	7.90	43	42	0.6	8,300	0.1	4	1,170	1,010	7.2	<0.05	134		83	600	12
27												185				
28	7.90	108	92		7,670			1,100	1,040	8.0	1.0	175		84	1,000/1,200	4
29												145				
30	7.70	117	78	<0.2	6,300		8	902	454	7.0	0.1	55		32	800/800	4
12/01/79												45				
02																<1

continued

Appendix B-1 (continued)

Date	pH	SS mg/l	VSS mg/l	$S^=$ mg/l	TDS mg/l	Phenol mg/l	O&G mg/l	TKN as N mg/l	NH_3 as N mg/l	CN_T mg/l	CN_A mg/l	TOC mg/l	Total P as P mg/l	BOD mg/l	Color APHA Units	SCN mg/l
Activated Sludge II - Clarifier Effluent (continued)																
12/03/79		37	27			<0.1	25	423	400	2.9	0.4	35	1.4	5		<1
04												37				
05		34	26			0.2	41	375	350	2.1	<0.05	37	1.0	7		<1
06												42				
07		23	12			<0.1	26	364	356	1.8	1.0	35	1.2	3		<1
08												44				
09												27				
10		28	14			0.1	15	336	308	1.9	<0.05	22	0.2	5		<1
11												16				
12		20	16					314	277	2.1	<0.05	8	0.3		200/300	<1
13												16				
14		24	18	7.0		2.8	5	313	280	1.4	<0.05	7	0.5	5	150/200	<1
15												9				
16												5				
17		15	11			<0.1	4	532	263	4.5	2.5	13	<0.2	3	100/150	<1
18												3				
19		22	15					288	218	1.2	<0.05	6	<0.2	3	50/100	<1
20												4				
21		20	13	<0.2		0.8	3	241	213	1.1	<0.05	6	1.2	3	50/100	<1
22												2				
23												22				
24		12	11			1.2	5	185	140	0.5	<0.05	6		3	100/150	<1
25												2				
26		17	16			0.1		182	174	1.1	<0.05	11	2.3	1		<1
27												11				
28		48	21	<0.2		0.1	8	186	174	0.6	<0.05	9		5	150/200	<1
29												12				
30												5				
31		34	12			0.6		187	162	0.1	<0.05	8				<1
1/01/80												8				
02		35	29	<0.2	2,760	<0.1	5	185	162	1.7	<0.05	5	1.9	2	200/300	1
03												43				
04		50	20					218	179	3.5	<0.05	30	2.5	6		2
05												24				
06												21				
07		28	18			6.2		95	78	3.4	1.9	15	1.5	25		2
08												18				
09		34	18						39	3.1	<0.05	30	0.9	15		<1
10												22				
11		25	13	A		<0.1	8	22	17	2.1	0.2	21	1.3	5	100/150	<1

A: Denotes Interference

continued

Appendix B-1 (continued)

Date	pH	SS mg/l	VSS mg/l	$S^=$ mg/l	TDS mg/l	Phenol mg/l	O&G mg/l	TKN as N mg/l	NH_3 as N mg/l	CN_T mg/l	CN_A mg/l	TOC mg/l	Total P as P mg/l	BOD mg/l	Color APHA Units	SCN mg/l
Activated-Sludge II - Clarifier Effluent (continued)																
1/12/80												24				
13												26				
14		47	16			<0.1	8	20	17	6.1	0.8	26	1.9	4		<1
15												32				
16		56	30					20	17	7.0	2.6	36	2.7	5		<1
17												38				
18		66	41	A		<0.1	11	11	8	5.1	<0.05	51	3.7	7	150/200	<1
19												59				
20												62				
21		66	40			0.6	8	10	2	5.4	<0.05	67	3.1	6		<1
22												72				
23		51	32					9	1	4.8	<0.05	104	3.6	6		<1
24												70				
25		59	29	A		0.1	6	9	3	7.1	<0.05	57	3.8	7	400/600	<1
26												40				
27												58				
28		93	40			<0.1	3	8	6	8.9	<0.05	60	3.3	6		<1
29												115				
30		253	145			<0.1		73	45	30.2	<0.05	110	4.6	20		65
31												127				
2/01/80		306	173	A		<0.1	5	50	22	8.9	5.10	114	2.5	50	500/600	<1
02		337	167	A		<0.1	6	28	17	9.0	<0.05	120	2.3	39	400/600	<1
03		230	142	A		<0.1	12	42	22	9.7	<0.05	154	2.4	42	400/500	<1
04		275	150	A		0.1	12	39	22	8.1	<0.05	104	2.0	41	300/350	<1
05		340	220	A		0.4	21	31	11	8.6	<0.05	158	2.6	87	300/400	<1
06		240	230	A		0.7	12	17	1	8.0	0.4	104	4.2	56	400/500	<1
07		150	95	A		0.9	14	17	2	7.0	<0.05	114	4.0	24	500/600	<1

A: Denotes Interference

Appendix B-1 (continued)

ADDITIONAL ANALYTICAL DATA

Date	System feed mg/l	Activated sludge I -clarifier effluent mg/l	Activated sludge II -clarifier effluent mg/l
Nitrite/Nitrate			
10/24/79	2	-	-
12/17/79	2	1	4
12/19/79	<1	1	5
1/2/80	<1	1	8
1/9/80	5	2	-
1/18/80	-	1	205
1/21/80	-	2	180
1/23/80	-	2	160
2/1/80	-	1	215
2/2/80	-	1	195
2/4/80	-	1	170
2/5/80	-	2	170
2/6/80	-	1	185
2/7/80	-	2	190
Sulfate			
1/2/80	1,600	1,980	700
Chloride			
1/2/80	4,500	3,700	1,200
Sodium			
10/24/79	1,700	-	-
10/25/79	2,400	1,800	-
Potassium			
10/24/79	5.3	-	-
10/25/79	5.4	5.0	-
Calcium			
10/24/79	1.9	-	-
10/25/79	2.5	1.5	-
Magnesium			
10/24/79	5.2	-	-
10/25/79	3.4	5.8	-
Manganese			
10/24/79	0.4	-	-
10/25/79	<0.1	0.4	-

Appendix B-1 (continued)

ADDITIONAL ANALYTICAL DATA

Date	System feed mg/l	Activated sludge I -clarifier effluent mg/l	Activated sludge II -clarifier effluent mg/l
Iron			
10/24/79	7.5	-	-
10/25/79	2.9	2.2	-
Alkalinity as $CaCO_3$			
10/24/79	535	-	-
11/5/79	210	104	-
12/24/79	340	680	250
12/26/79	220	770	330
12/28/79	220	760	360
1/2/80	430	1,000	440

APPENDIX B-2

Contract 68-02-2685
Technical Directive 106

SATS
PRIORITY POLLUTANT ANALYSIS FOR COKE PLANT WASTEWATER TREATMENT

Final Report, Technical Directive 106
Period Covered: October 12, 1979 through May 15, 1980

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SATS
PRIORITY POLLUTANT ANALYSIS FOR COKE PLANT WASTEWATER TREATMENT

The purpose of this contract was to obtain and analyze wastewater samples from EPA's Mobile Wastewater Treatment System (MWWTS) project being conducted at the Wheeling-Pittsburgh Steel Company in Follansbee, West Virginia.

A total of 13 samples plus appropriate blanks were collected during February 1980. The priority pollutant sampling was conducted by Davis and Floyd Engineers, Inc., Greenwood, South Carolina, under subcontract to Southern Research Institute. The samples were shipped to SoRI, Birmingham, Alabama, for analysis of priority pollutants and to Davis and Floyd for traditionals. Sampling was conducted in accordance with the EPA screening protocol.

During the sampling operation, the MWWTS was treating coke plant waste in a two-stage biological system followed by an activated carbon adsorption unit. A schematic of the MWWTS is given in Figure 1. A site reference sampling code for the samples taken is given in Table 1. For additional details relative to the operation of the MWWTS during the sampling period (from February 4, 1980, to February 7, 1980), refer to the sample crew log sheets in the appendix.

The results of the analyses are given in Tables 2 through 4. Quality control/quality assurance data from analyses of duplicate and spiked samples are given in Tables 5 through 14.

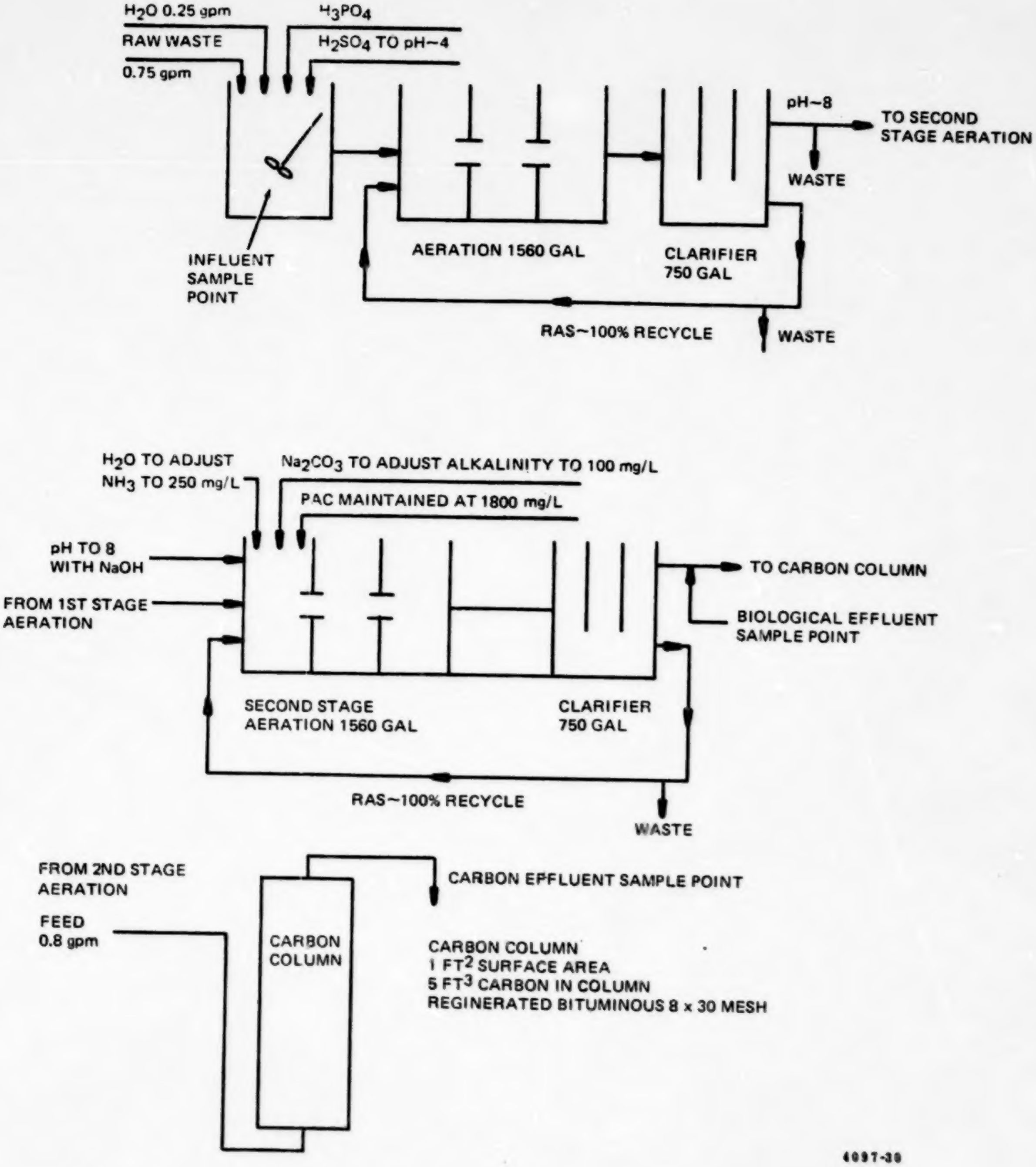
H2O 0.25 gpm
RAW WASTE
0.75 gpm
H3PO4
H2SO4 TO pH~4
pH~8
TO SECOND STAGE AERATION
WASTE
INFLUENT SAMPLE POINT
AERATION 1560 GAL
CLARIFIER 750 GAL
RAS~100% RECYCLE
WASTE
H2O TO ADJUST NH3 TO 250 mg/L
Na2CO3 TO ADJUST ALKALINITY TO 100 mg/L
PAC MAINTAINED AT 1800 mg/L
pH TO 8 WITH NaOH
FROM 1ST STAGE AERATION
TO CARBON COLUMN
BIOLOGICAL EFFLUENT SAMPLE POINT
SECOND STAGE AERATION 1560 GAL
CLARIFIER 750 GAL
RAS~100% RECYCLE
WASTE
FROM 2ND STAGE AERATION
FEED 0.8 gpm
CARBON COLUMN
CARBON EFFLUENT SAMPLE POINT
CARBON COLUMN
1 FT2 SURFACE AREA
5 FT3 CARBON IN COLUMN
REGINERATED BITUMINOUS 8 x 30 MESH
4097-30

Figure 1. EPA pilot plant schematic Follansbee, W.V.

TABLE 1. SAMPLE REFERENCE CODE

Date	Biological influent	Biological effluent	Carbon effluent	Raw water[2]
2-4-80	01678[1]	06179[1]	06180[1]	06181[1]
2-5-80	06182	06183	06184	-
2-6-80	06185	06186	06187	-
2-7-80	06188	06189	06190	-

1. Grab samples

2. Tap water used for dilution

TABLE 2. VERIFICATION REPORT FOR TRADITIONALS

Traditionals	Activated sludge system influent			Activated sludge system effluent			Carbon column effluent		
	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80
Total cyanide	195	220	170	7.9	7.8	6.6	8.0	6.6	5.0
Cyanide amenable to chlorination	160	198	150	<0.1	<0.1	<0.1	0.4	<0.1	<0.1
Thiocyanate	462	486	466	2.9	2.0	1.7	1.9	1.5	0.9
Phenols	1000	900	750	0.500	0.011	0.030	0.006	0.005	0.004
Sulfide	52	50	16	36	11	0.6	1.2	0.6	0.6
Oil & Grease	10.8	24.0	22.0	2.4	2.1	7.3	2.0	7.5	4.4
Ammonia as N	278	270	294	46	8.9	2.7	11.9	3.9	2.1
Nitrate as N	<0.05	<0.05	<0.05	79	122	200	87	138	178
Total suspended solids	68	64	52	300	260	240	54	100	60
Color	450	2500	400	1000	500	450	200	100	100

Notes: 1. All results in mg/L except Color which is reported in APHA units.

2. (<) represents the lower limit of detection for the method and/or instrument.

3. Total phenol (4-AAP method).

4. It is recognized that the influent cyanide concentrations shown in this table are well above the influent values shown in Appendix B-1. The reason for this discrepancy could not be determined.

TABLE 3. VERIFICATION REPORT FOR METALS

Metals	Raw water	Activated sludge system influent			Activated sludge system effluent			Carbon column effluent		
	2/04/80	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80
Ag	<10	18	18	14	12	12	12	12	12	12
As	<10	350	340	390	86	95	140	86	82	130
Be	<10	<10	<10	<10	<10	<10	<10	<10	<10	<10
Cd	<15	<15	<15	<15	<15	<15	<15	<15	<15	<15
Cr	<10	21	24	25	<10	<10	10	<10	<10	<10
Cu	<50	71	71	57	<50	<50	<50	<50	<50	<50
Hg	<0.1	7.8	3.7	5.6	3.8	2.1	1.2	0.2	0.6	-
Ni	<100	<100	<100	<100	<100	<100	<100	<100	<100	<100
Pb	<10	<10	<10	<10	<10	<10	<10	<10	<10	<10
Sb	<10	260	120	120	<20	<20	<20	<20	<20	<20
Se	<10	1400	1400	1320	330	350	550	250	290	420
Tl	<10	<10	<10	<10	<10	<10	<10	<10	<10	<10
Zn	630	480	460	390	1400	1200	920	300	560	330

Notes: 1. All results in µg/L.

2. (<) represents the lower limit of detection for the method and/or instrument.

TABLE 4. GC/MS VERIFICATION REPORT FOR ORGANICS

Compound	Raw Water	Activated sludge system influent			Activated sludge system effluent			Carbon column effluent		
	2/04/80	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80	2/05/80	2/06/80	2/07/80
Methylene chloride	15	589	548	489	<10	<10	<10	22	<10	<10
Toluene	<10	722	542	556	<10	<10	14	<10	<10	15
Chloroform	ND	<10	<10	<10	<10	<10	25	<10	<10	31
Trichloroethylene	10	ND	ND	ND	ND	ND	ND	ND	ND	ND
Bromodichloromethane	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Dibromochloromethane	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Benzene	15	9,820	6,100	6,210	ND	ND	ND	ND	ND	ND
Naphthalene	ND	964	760	1,160	<10	ND	ND	ND	ND	ND
Acenaphthylene	ND	59	40	57	ND	ND	ND	ND	ND	ND
Fluorene	ND	48	33	34	ND	ND	ND	ND	ND	ND
Anthracene/Phenanthrene	ND	562	454	582	<10	<10	<10	<10	<10	ND
Fluoranthene	ND	333	311	410	<10	<10	<10	ND	ND	ND
Pyrene	ND	212	196	265	<10	<10	<10	ND	ND	ND
Chrysene/ Benzo(a)anthracene	ND	84	86	100	ND	ND	<10	ND	<10	ND
Benzo(b)fluoranthene	ND	16	12	13	ND	ND	<10	ND	ND	ND
Benzo(a)pyrene	ND	<10	ND	ND	ND	ND	<10	ND	ND	ND
2-Chlorophenol	ND	985	1,660	1,170	ND	ND	ND	ND	ND	ND
2-Nitrophenol	ND	29	<25	<25	<25	ND	ND	<25	ND	ND
Phenol	ND	65,400	65,200	55,800	ND	ND	<25	ND	ND	ND
2,4-Dimethylphenol	ND	6,810	6,970	7,600	ND	ND	ND	ND	ND	ND

Notes: 1. All results are reported as μg/L unless indicated.

2. ND = Not Detected

QUALITY ASSURANCE SUMMARY

TABLE 5. QUALITY ASSURANCE SUMMARY FOR TRADITIONALS PERCENT RECOVERY

Traditionals	Activated sludge system influent	Activated sludge system effluent	Carbon column effluent
Total cyanide	90	88	104
Thiocyanate	100	84	104
Phenols	80	88	100
Ammonia as N	88	98	101
Nitrate as N	95	95	105

TABLE 6. QUALITY ASSURANCE SUMMARY FOR METALS PERCENT RECOVERY

	Raw Water				Carbon Column Effluent				Activated Sludge System Effluent				Activated Sludge System Influent			
Metals	Spiked	Unspiked	Amount added	Recovery, %	Spiked	Unspiked	Amount added	Recovery, %	Spiked	Unspiked	Amount added	Recovery, %	Spiked	Unspiked	Amount added	Recovery %
Ag	24	<10	20	118	31	13	20	94	31	13	20	94	31	13	20	94
As	97	<10	100	97	245	148	100	97	258	163	100	95	812	377	400	109
Hg	0.95	<0.1	1.00	95	1.56	0.66	1.00	90	1.96	1.40	1.00	56	7.9	1.65	10.00	63
Sb	97	<5	100	97	94	<15	100	94	97	20	100	97	210	100	100	110
Se	303	<10	300	101	850	504	300	115	888	610	300	93	2603	1282	1500	88
Zn	1094	613	500	96	780	320	500	92	1802	884	1000	92	832	347	500	97

Notes: 1. All results in μg/L.

2. $\% \text{ Recovery} = 100 \left(\frac{\text{Spiked} - \text{Unspiked}}{\text{Amount added}} \right)$

3. All values are averages of duplicate samples.

4. (<) represents the lower limit of detection for the method and/or instruments.

TABLE 7. QUALITY ASSURANCE SUMMARY FOR METALS REPLICATE ANALYSES

Metal	Activated sludge system influent			Activated sludge system effluent			Carbon column effluent		
	No. 1	No. 2	No. 3	No. 1	No. 2	No. 3	No. 1	No. 2	No. 3
Ag	18	13	13	12	13	13	12	13	13
As	430	410	344	110	148	178	110	148	147
Hg	7.8	1.8	1.5	1.2	1.2	1.8	0.5	0.8	0.5
Sb	150	100	NA	<20	<20	<20	<20	<20	<20
Sc	1400	1333	1230	440	571	648	270	528	479
Zn	480	351	343	1000	884	884	360	316	324

1. All results in μg/L.

TABLE 8. GC/MS VERIFICATION QUALITY ASSURANCE SUMMARY FOR ORGANICS ACTIVATED SLUDGE SYSTEM INFLUENT

Compound	Concentration found in unspiked samples, µg/L		Concentration found in spiked samples, µg/L	Concentration spiked, µg/L	Average recovery, %
	No. 1	No. 2			
1,3-Dichlorobenzene	ND	ND	218	100	218
1,4-Dichlorobenzene	ND	ND	79	100	79
1,2-Dichlorobenzene	ND	ND	46	100	46
Hexachloroethane	ND	ND	53	100	53
Bis(2-chloroisopropyl)ether	ND	ND	61	100	61
Nitrobenzene	ND	ND	<10	100	<10
Hexachlorobutadiene	ND	ND	63	100	63
Naphthalene	1160	282	393	100	111
1,2,3-Trichlorobenzene	ND	ND	63	100	63
Bis(2-chloroethoxy)methane	ND	ND	78	100	78
N-nitroso-di-n-propyl amine	ND	ND	20	100	20
Isophrone	ND	<10	11	100	<10
2-Chloronaphthalene	ND	ND	58	100	58
Acenaphthylene	57	18	86	100	68
Acenphthene	ND	<10	73	100	66
2,6-Dinitrotoluene	ND	ND	140	100	140
Fluorene	34	17	75	100	58
2,4-Dinitrotoluene	ND	ND	80	100	80
1,2-Diphenylhydrazine	ND	ND	105	200	52
N-nitrosodiphenylamine	ND	ND	108	100	108
Anthracene/Phenanthrene	582	214	353	200	70
Fluoranthene	410	206	309	100	103
Pyrene	265	150	239	100	89
Benzidine	ND	ND	77	200	39
Benzo(a)anthracene/Chrysene	100	141	332	200	96
Benzo(b)fluoranthene	13	114	242	100	128
Benzo(a)pyrene	ND	114	211	100	97
Indeno(1,2,3-cd)pyrene	ND	48	333	200	143
Dibenzo(ah)anthracene	ND	37	182	100	144
Benzo(ghi)perylene	ND	28	167	100	139
2-Chlorophenol	1170	684	827	200	71
2-Nitrophenol	<25	<25	201	200	98
Phenol	55800	4840	6040	200	AI
2,4-Dimethylphenol	7600	1230	1470	200	120
2,4-Dichlorophenol	ND	<25	88	200	43
2,4,6-Trichlorophenol	ND	<25	221	200	109
4-Chloro-3-methylphenol	ND	<25	204	200	98
2,4-Dinitrophenol	ND	<25	1360	1000	136
4,6-Dinitro-o-cresol	ND	<25	2190	1500	145
Pentachlorophenol	ND	<25	255	200	126

Note: **ND = Not Detected**

AI = Analytical Interference

TABLE 9. GC/MS VERIFICATION
QUALITY ASSURANCE SUMMARY FOR ORGANICS
ACTIVATED SLUDGE SYSTEM EFFLUENT

Compound	LEVEL I						LEVEL II					
	Concentration found in unspiked samples, µg/L			Concentration found in spiked sample, µg/L	Concentration spiked, µg/L	Average recovery, %	Concentration found in unspiked samples, µg/L			Concentration found in spiked sample, µg/L	Concentration spiked, µg/L	Average recovery, %
	No. 1	No. 2	No. 3				No. 1	No. 2	No. 3			
1,3-Dichlorobenzene	ND	ND	ND	36	20	182	ND	ND	ND	38	50	76
1,4-Dichlorobenzene	ND	ND	ND	16	20	79	ND	ND	ND	16	50	33
1,2-Dichlorobenzene	ND	ND	ND	<10	20	33	ND	ND	ND	23	50	46
Hexachloroethane	ND	ND	ND	<10	20	43	ND	ND	ND	21	50	41
Bis(2-chloroiospropyl)ether	ND	ND	ND	21	20	105	ND	ND	ND	38	50	76
Nitrobenzene	ND	ND	ND	22	20	108	ND	ND	ND	28	50	57
Hexachlorobutadiene	ND	ND	ND	10	20	48	ND	ND	ND	24	50	49
Naphthalene	ND	ND	ND	13	20	63	ND	ND	ND	27	50	53
1,2,3-Trichlorobenzene	ND	ND	ND	14	20	68	ND	ND	ND	29	50	58
Bis(2-chloroethoxy)methane	ND	ND	ND	20	20	99	ND	ND	ND	39	50	78
N-nitroso-di-n-propyl amine	ND	ND	ND	24	20	122	ND	ND	ND	39	50	77
Isophorone	ND	ND	ND	16	20	79	ND	ND	ND	33	50	66
2-Chloronaphthalene	ND	ND	ND	14	20	69	ND	ND	ND	28	50	56
Acenaphthylene	ND	ND	ND	13	20	64	ND	ND	ND	30	50	60
Acenaphthene	ND	ND	ND	15	20	77	ND	ND	ND	32	50	65
2,6-Dinitrotoluene	ND	ND	ND	23	20	113	ND	ND	ND	58	50	117
Fluorene	ND	ND	ND	14	20	69	ND	ND	ND	28	50	57
2,4-Dinitrotoluene	ND	ND	ND	25	20	124	ND	ND	ND	51	50	103
1,2-Diphenylhydrazine	ND	ND	ND	22	40	55	ND	ND	ND	48	100	48
N-nitrosodiphenylamine	ND	ND	ND	20	20	102	ND	ND	ND	39	50	78
Anthracene/Phenanthrene	<10	<10	<10	19	40	36	<10	<10	<10	45	100	40
Fluoranthene	<10	<10	<10	13	20	30	<10	<10	<10	31	50	48
Pyrene	<10	<10	<10	14	20	31	<10	<10	<10	33	50	50
Benzidine	ND	ND	ND	ND	40	<10	ND	ND	ND	47	100	47
Benzo(a)anthracene/Chyrsene	ND	ND	<10	14	40	27	ND	ND	<10	69	100	66
Benzo(b)fluoranthene	ND	ND	<10	<10	20	14	ND	ND	<10	47	50	87
Benzo(a)pyrene	ND	ND	12	<10	20	<10	ND	ND	12	46	50	84
Indeno(1,2,3-cd)pyrene	ND	ND	ND	ND	40	<10	ND	ND	ND	79	100	79
Dibenzo(ah)anthracene	ND	ND	ND	ND	20	<10	ND	ND	ND	40	50	80
Benzo(ghi)perylene	ND	ND	ND	ND	20	<10	ND	ND	ND	43	50	86
2-Chlorophenol	ND	ND	ND	26	40	64	ND	ND	ND	62	100	62
2-Nitrophenol	ND	ND	ND	27	40	69	ND	ND	ND	78	100	78
Phenol	<25	<25	<25	<25	40	32	<25	<25	<25	41	100	36
2,4-Dimethylphenol	ND	ND	ND	<25	40	51	ND	ND	ND	61	100	61
2,4-Dichlorophenol	ND	ND	ND	31	40	77	ND	ND	ND	79	100	79
2,4,6-Trichlorophenol	ND	ND	ND	29	40	73	ND	ND	ND	93	100	93
4,Chloro-3-methylphenol	ND	ND	ND	32	40	81	ND	ND	ND	91	100	91
2,4-Dinitrophenol	ND	ND	ND	159	200	80	ND	ND	ND	611	500	122
4,6-Dinitro-o-cresol	ND	ND	ND	613	500	123	ND	ND	ND	1250	1000	125
Pentachlorophenol	ND	ND	ND	33	40	83	ND	ND	ND	119	100	119

Note: ND = Not Detected

TABLE 10. GC/MS VERIFICATION
QUALITY ASSURANCE SUMMARY FOR ORGANICS
CARBON COLUMN EFFLUENT

Compound	Concentration found in unspiked samples, µg/L		Concentration found in spiked samples, µg/L	Concentration spiked, µg/L	Average recovery, %
	No. 1	No. 2			
1,3-Dichlorobenzene	ND	ND	35	20	174
1,4-Dichlorobenzene	ND	ND	18	20	91
1,2-Dichlorobenzene	ND	ND	11	20	56
Hexachloroethane	ND	ND	<10	20	46
Bis(2-chloroisopropyl)ether	ND	ND	28	20	140
Nitrobenzene	ND	ND	<10	20	<10
Hexachlorobutadiene	ND	ND	12	20	60
Naphthalene	ND	ND	15	20	74
1,2,3-Trichlorobenzene	ND	ND	16	20	80
Bis(2-chloroethoxy)methane	ND	ND	23	20	114
N-nitroso-di-N-propyl amine	ND	ND	22	20	111
Isophrone	ND	ND	17	20	83
2-chloronaphthalene	ND	ND	20	20	99
Acenaphthylene	ND	ND	14	20	68
Acenaphthene	ND	ND	20	20	99
2,6-Dinitrotoluene	ND	ND	18	20	91
Fluorene	ND	ND	19	20	94
2,4-Dinitrotoluene	ND	ND	22	20	111
1,2-Diphenylthydrazine	ND	ND	35	40	87
N-nitrosodiphenylamine	ND	ND	21	20	104
Anthracene/Phenanthrene	ND	ND	37	40	93
Fluoranthene	ND	ND	21	20	107
Pyrene	ND	ND	28	20	139
Benzidine	ND	ND	<10	40	<10
Benzo(a)anthracene/chrysene	ND	ND	48	40	120
Benzo(b)fluoranthene	ND	ND	23	20	113
Benzo(a)pyrene	ND	ND	22	20	111
Indeno(1,2,3-cd)pyrene	ND	ND	25	40	64
Dibenzo(ah)anthracene	ND	ND	<10	20	28
Benzo(ghi)perylene	ND	ND	19	20	96
2-Chlorophenol	ND	ND	<25	40	40
2-Nitrophenol	ND	ND	<25	40	45
Phenol	ND	ND	<25	40	24
2,4-Dimethylphenol	ND	ND	<25	40	<10
2,4-Dichlorophenol	ND	ND	<25	40	43
2,4,6-Trichlorophenol	ND	ND	<25	40	42
4-Chloro-3-methylphenol	ND	ND	<25	40	42
2,4-Dinitrophenol	ND	ND	92	200	46
4,6-Dinitro-o-cresol	ND	ND	350	500	70
Pentachlorophenol	ND	ND	<25	40	47

Note: ND = Not Detected

TABLE 11. GC/MS VERIFICATION QUALITY ASSURANCE SUMMARY FOR VOA PERCENT RECOVERY ACTIVATED SLUDGE SYSTEM INFLUENT

Compound	Concentration found in unspiked sample, μg/L	Concentration found in spiked sample, μg/L	Concentration spiked, μg/L	Average recovery, %
Methylene chloride	1150	1390	300	79
Toluene	836	1060	300	74
Chloroform	5.9	317	300	104
Bromodichloromethane	ND	298	300	99
Dibromochloromethane	ND	228	300	76
Trichloroethylene	ND	303	300	101

TABLE 12. GC/MS VERIFICATION
QUALITY ASSURANCE SUMMARY FOR VOA
PERCENT RECOVERY
CARBON COLUMN EFFLUENT

	LEVEL I				LEVEL II			
Compound	Concentration found in unspiked samples, µg/L	Concentration found in spiked samples, µg/L	Concentration spiked µg/L	Average recovery, %	Concentration found in unspiked samples µg/L	Concentration found in spiked sample, µg/L	Concentration spiked µg/L	Average recovery, %
Methylene chloride	22	38	20	80	<10	130	100	123
Trichlorofluoromethane	ND	19	20	94	ND	134	100	134
1,1-Dichloroethylene	ND	24	20	121	ND	150	100	150
Trans-1,2-dichloroethylene	ND	21	20	107	ND	119	100	119
1,1-Dichloroethane	ND	21	20	107	ND	126	100	126
Chloroform	<10	24	20	100	31	116	100	85
1,2-Dichloroethane	ND	19	20	94	ND	120	100	120
1,1,1-Trichloroethane	ND	21	20	106	ND	119	100	119
1,1,2-Trichloroethane	ND	20	20	102	ND	106	100	106
Carbon tetrachloride	ND	19	20	94	ND	132	100	132
1,2-Dichloropropane	ND	24	20	119	ND	102	100	102
Trans-1,3-dichloropropene	ND	24	20	118	ND	112	100	112
Bromodichloromethane	ND	16	20	80	ND	102	100	102
Dibromochloromethane	ND	20	20	99	ND	111	100	111
Chlorobenzene	ND	20	20	99	ND	100	100	100
Trichloroethylene	ND	21	20	107	ND	125	100	125
Benzene	ND	24	20	118	ND	116	100	116
Ethylbenzene	ND	21	20	106	ND	100	100	100
Toluene	<10	22	20	90	15	109	100	94
Bromoform	ND	21	20	106	ND	112	100	112
1,1,2,2-Tetrachloroethene	ND	20	20	102	ND	114	100	114
1,1,2,2-Tetrachloroethane	ND	26	20	129	ND	118	100	118

Note: ND = Not Detected

TABLE 13. GC/MS VERIFICATION QUALITY ASSURANCE SUMMARY REPLICATE ANALYSES

	Activated sludge system influent		Activated sludge system effluent		Carbon column effluent	
	No. 1	No. 2	No. 1	No. 2	No. 1	No. 2
Methylene chloride	520	575	3.9	4.0	4.1	4.7
Toluene	468	616	3.1	2.6	3.2	3.0
Chloroform	4.8	*	4.5	4.3	ND	ND
Benzene	6100	*	ND	ND	ND	ND

* Concentrations outside working linear range.
All results in µg/L.
ND = Not Detected

TABLE 14. GC/MS VERIFICATION
SPIKE RECOVERY DATA
PERCENT RECOVERY

Compound	Carbon Effluent	Biological Influent
Methylene chloride	102	74
Toluene	92	80
Chloroform	92	102
Bromodichloromethane	91	104
Dibromochloromethane	104	86
Trichloroethylene	116	100
Benzene	117	-

Note: $$\% \text{ Recovery} = \frac{\text{spiked - unspiked}}{\text{amount spiked}} \times 100$$

APPENDIX

SAMPLE CREW LOG SHEETS

Davis & Floyd Engineers, Inc.

INTER-OFFICE MEMO

TO: File

FROM: Gary Anderson

DATE: 2-15-80

SUBJECT: Priority Pollutant Sampling
Wheeling Pittsburgh Steel Company
Follansbee Coke Plant
Follansbee, West Virginia

JOB: 3076

On February 4, 1980, Davis & Floyd arrived at Wheeling-Pittsburgh Steel Company, Follansbee Coke Plant at 0730 hours. A brief introduction to the pilot plant system was given to us by Dick Osentowski of Rexnord. A schematic diagram of the pilot plant operation was obtained. The three previously determined sampling points were located (biological influent, biological effluent, and carbon column effluent). The carbon effluent sample was taken directly from the final carbon column.

At 1000 hours, each automatic sampler was prepared and placed at each sample point. One liter of triple distilled blank water was pumped through each sampler and discarded. A one gallon sample of blank water was pumped through and collected from each sampler, set up and prepared for shipment to Southern Research Institute. ISCO sampler No. B-1231-59 was set up on the biological influent. This sampler was equipped with 7 feet of 1/4 inch Teflon tubing and had a 1 foot head. This sample point was the mixed influent tank. The biological effluent ISCO sampler, No. B-1211-46, was set up. This sampler had 11 feet of 1/4 inch Teflon tubing with 1 foot of head. The carbon column effluent ISCO sampler, No. B-1066-17, was set up. The Teflon tubing from this sampler was attached to a sample port on the carbon column. This sampler had 7 feet of 1/4 inch Teflon tubing and no head. The carbon effluent sampler was under constant pressure between 15 PSI and 30 PSI throughout the sampling period. One gallon grab samples and VOA samples were obtained from the biological influent, biological effluent, and carbon effluent at 1045 hours and prepared for shipment to SRI 4. Like samples were also taken of dilution water which Rexnord used to dilute the raw wastewater. These samples were taken from a faucet in the pilot plant. Grab sample numbers were:

06178 - Biological influent
06179 - Biological effluent
06180 - Carbon effluent
06181 - For dilution water

Composite sampling started at 1200 hours. At each sampling point (Numbers 06182, 06183, 06184), a 200 ml aliquot of sample was collected every 30 minutes for 24 hours. Two sets of VOA samples were collected daily at each sample point. These were taken at approximately 2 hours after sampling started and approximately 2 hours prior to the end of the 24 hour composite period unless otherwise noted. The pilot plant was operating normally and flow was one GPM through the system. No detectable chlorine was present in the system. The 24 hour composite sample and VOA's from each sampling point were shipped daily to Southern Research Institute, Birmingham, Alabama via U. S. Air Express.

Davis & Floyd, Inc., Greenwood, South Carolina shipped the following samples daily:

2 liter CN, SCN (composite sample from 2 1/2 gal. composite)
1 liter Phenol (composite sample)
1 liter NH_3 (composite sample)
1 liter NO_3, Color, TSS (composite sample)
1 liter Oil & Grease (grab sample)
100 ml sulfide (grab sample)

These samples were preserved according to protocol and shipped:

Flight 327 Pittsburgh to Atlanta
Flight 784 Atlanta to Greenville-Spartanburg

The samplers were checked at 1630 hours. During this check, the carbon effluent sampler was noted to have malfunctioned due to pressure. The pressure from the carbon unit was adjusted and the sampler reset. VOA samples were collected at 1630 hours with the biological influent sample being amber colored. The biological effluent was black and the carbon effluent was light gray.

The following is pilot plant analysis data (2-4-80) obtained from Renord personnel:

1. Influent ammonia - 272 mg/l
2. Primary clarifiery effluent ammonia - 703 mg/l
3. Secondary clarifier effluent ammonia - 0.6 mg/l
4. Carbon effluent sulfide - <1 mg/l
5. MLSS Primary aeration - 7,300 mg/l
6. TSS Secondary clarifier effluent - 187 mg/l
7. Flow: First stage effluent - .25 gpm
   Dilution water - .75 gpm

February 5, 1980

Davis & Floyd arrived at the plant site at 0945 hours and found the pilot plant operating normally. VOAs were collected and preserved. The biological effluent and carbon effluent contained noticeable suspended solids. The color of these samples was the same as 2-4-80. The second 24 hour sampling period began at 1200 hours.

The following is pilot plant analysis data (2-5-80) obtained from Rexnord personnel:

1. Influent ammonia - 566 mg/l
2. Primary clarifier effluent ammonia - 569 mg/l
3. Secondary clarifier effluent ammonia - 0.8 mg/l
4. Carbon effluent sulfide - <1 mg/l
5. MLSS on primary aeration - 7,300 mg/l
6. TSS on secondary clarifier effluent - 220 mg/l
7. TSS on carbon effluent - 39 mg/l
8. Flow: First stage effluent - .44 gpm
   Dulution water - .56 gpm

February 6, 1980

Davis & Floyd arrived at the plant at 0900 hours. The pilot plant system was operating normally. VOAs were collected with the color of these samples same as previously indicated. Suspended solids were present in the biological effluent and the carbon effluent. The samplers for Davis & Floyd were poured from the 2 1/2 gallon composite jug and the Southern Research volume was approximately 2 gallons. Due to high concentration in biological influent and possible interferences during analysis, the biological influent sample was treated with cadmium carbonate.

Due to higher than normal suspended solids, Rexnord backwashed the carbon filter. The carbon effluent sampler was turned off at 1215 hours for the backwash. Rexnord explained that the carbon filter was accumulating thick black sludge causing the filter to clog. The carbon filter was backwashed with dilution water for 45 minutes. The carbon system was placed back on line at 1315 hours and carbon effluent sampler restarted.

The following is pilot plant analysis data (2-6-80) obtained from Rexnord personnel:

1. Influent ammonia - 404 mg/l
2. Primary clarifier effluent ammonia - 572 mg/l
3. Secondary clarifier effluent ammonia - 1.7 mg/l
4. Carbon effluent sulfide - <1 mg/l
5. MLSS primary aeration - 6,550 mg/l
6. MLSS secondary aeration - 6,900 mg/l
7. TSS primary effluent - 85 mg/l
8. TSS secondary effluent - 160 mg/l
9. TSS carbon effluent - 58 mg/l

February 7, 1980

Davis & Floyd arrived at the plant at 0900 hours. The pilot plant was operating normally. VOAs were collected from all sample points. The suspended solids had diminished considerably in the carbon effluent. Davis & Floyd and Southern Research samples collected and shipped per usual.

HGA/ptm

Appendix B-3.1 PILOT STUDY pH PROFILES (FIELD DATA)

Date	Time	Undiluted Waste to Neutralization Tank	Neutralization Tank		AS-1		AS-2		AS-2 Effluent
			Pump On	Pump Off	Cell 1	Cell 3	Cell 1	Cell 3	
11/24/79	0730	9.2	---	6.8	7.25	7.35	8.0	8.2	8.2
25	1100	8.7	---	---	7.3	7.75	7.8	7.8	7.9
26	0800	8.7	7.5	7.0	7.1	7.4	7.8	7.55	7.7
27	1145	8.8	---	---	7.3	7.5	7.9	7.8	7.8
28	1005	8.95	---	---	7.2	7.5	---	7.9	7.9
29	0815	8.8	7.6	6.4	7.3	7.4	7.6	7.6	7.75
30	0935	9.0	7.8	7.1	7.3	7.4	7.6	7.75	7.75
12/01/80	1200	8.9	7.7	6.8	7.65	7.6	7.85	7.95	7.95
02	0930	8.9	7.8	6.8	7.4	7.45	7.6	7.8	7.85
03	1100	8.8	7.9	6.7	7.5	7.4	7.5	7.65	7.7
04	1100	8.7	7.7	6.7	7.5	7.3	7.3	7.55	7.65
05	1000	9.2	7.8	6.5	7.5	7.4	7.2	7.5	7.5
06	1030	9.2	7.8	6.75	7.4	7.3	7.15	7.4	7.4
07	1000	9.3	7.7	6.8	7.4	7.2	7.1	7.35	7.5
08	1100	9.0	7.8	6.8	7.2	7.1	7.0	7.3	7.4
09	1045	8.8	8.1	7.0	7.2	7.1	7.35	7.6	7.6
10	1130	9.3	7.9	6.8	7.0	7.5	7.3	7.6	7.6
11	1000	9.0	7.8	6.3	6.9	6.9	7.2	7.4	7.4
12	0945	8.8	7.8	6.4	7.0	6.7	7.2	7.3	7.4
13	0930	8.8	7.7	6.4	6.3	6.2	7.3	7.5	7.6
14	1030	8.9	7.8	6.4	6.8	6.6	7.3	7.3	7.4
15	1015	8.9	7.8	6.4	6.7	6.6	7.2	7.2	7.2
16	0940	8.5	7.8	6.3	6.7	6.7	7.2	7.3	7.3
17	0945	8.7	7.9	6.3	7.0	7.0	7.3	7.3	7.3
18	--	8.9	7.9	6.4	7.2	7.2	7.4	7.3	7.4

(continued)

Appendix B-3.1. (continued)

Date	Time	Undiluted Waste to Neutralization Tank	Neutralization Tank		As-1		As-2		AS-2 Effluent
			Pump On	Pump Off	Cell 1	Cell 3	Cell 1	Cell 3	
19	1500	8.8	7.9	6.4	7.3	7.3	7.3	7.3	7.3
20	0820	8.5	7.1	6.1	7.2	7.3	7.1	7.0	7.0
21	0815	8.3	7.3	5.5	7.7	7.85	7.6	7.6	7.5
22	0700	8.8	7.5	6.5	7.6	7.7	7.6	7.7	7.7
23	0850	8.0	7.5	5.9	7.8	7.9	7.6	7.6	7.6
24	0715	8.5	---	---	7.85	7.9	7.6	7.6	7.6
25	0630	8.6	---	---	7.95	8.0	7.6	7.7	7.7
26	1010	4.8	---	---	8.0	8.1	7.8	7.85	7.8
27	0640	8.4	---	---	7.7	8.0	7.7	7.8	7.8
28	0655	4.7	6.3	6.6	7.7	8.0	7.65	7.8	7.8
29	0820	4.7	6.3	6.85	7.75	7.7	7.8	7.8	7.8
30	0750	9.3	7.4	6.2	7.8	8.0	7.8	7.8	7.8
31	0710	8.3	---	---	7.8	8.0	7.9	7.9	7.85
1/01/80	0815	9.2	7.9	5.4	7.9	8.1	7.9	7.9	7.9
02	0645	9.2	---	5.2	7.9	8.1	7.9	7.9	7.9
03	0645	9.0	7.6	5.3	7.95	8.1	7.9	7.9	7.9
04	0700	9.0	---	---	8.05	8.3	7.7	7.5	7.55
05	0700	9.4	---	---	8.0	8.3	6.5	6.4	6.6
06	0710	9.1	7.0	4.5	7.9	8.25	6.9	6.1	6.2
07	0600	9.1	6.5	4.0	7.85	8.25	7.3	6.2	6.1
08	0720	9.0	---	---	7.9	8.2	7.9	6.9	6.5
09	0750	9.25	7.8	4.5	7.8	8.1	7.6	7.6	7.5
10	0730	9.0	7.0	4.3	7.8	8.0	7.5	7.5	7.4
11	0720	9.1	7.8	4.3	7.8	8.1	7.55	7.5	7.3

(continued)

Appendix B-3.1. (continued)

Date	Time	Undiluted Waste to Neutralization Tank	Neutralization Tank		AS-1		AS-2		AS-2 Effluent
			Pump On	Pump Off	Cell 1	Cell 3	Cell 1	Cell 3	
12	0740	9.2	7.6	4.3	7.85	8.1	7.65	7.5	7.4
13	0710	9.3	7.8	4.1	7.85	8.1	7.75	7.6	7.55
14	0720	9.1	---	---	7.8	8.0	7.7	7.0	7.0
15	0645	8.4	6.8	4.0	8.0	8.2	7.7	7.4	7.4
16	0900	8.2	7.0	3.9	8.0	8.4	7.4	7.3	7.2
17	0810	8.7	6.5	3.9	7.8	8.1	7.8	6.3	6.8
18	0700	9.8	---	---	8.1	8.5	8.4	7.4	7.4
19	0900	8.6	---	---	8.2	8.6	8.4	8.2	8.2
20	0900	9.4	7.4	3.9	8.4	8.6	8.6	8.6	8.5
21	0840	10.2	---	---	8.3	8.6	8.4	8.6	8.5
22	1000	9.9	7.6	3.9	8.2	8.5	7.8	8.4	8.3
23	1030	9.8	---	---	8.3	8.4	7.9	8.0	7.9
24	1515	9.9	---	---	7.8	8.4	7.4	7.6	7.2
25	0030	10.1	---	---	8.2	8.3	8.1	8.3	8.2
26	0930	10.0	8.0	4.0	8.1	8.2	8.3	8.1	8.0
27	0900	10.3	---	---	8.3	8.3	8.2	8.3	8.2
28	0915	10.2	---	---	8.2	8.2	8.4	7.8	7.8
29	0810	9.6	7.5	3.8	7.6	7.8	8.4	7.4	7.4
30	0845	9.7	4.6	4.1	7.7	8.2	8.0	8.0	7.4
31	0825	9.5	4.6	4.1	7.7	8.0	8.1	7.9	7.8
2/01/80	0915	9.2	4.6	4.0	7.7	8.0	8.2	7.3	7.3
02	0930	9.2	4.5	4.0	7.5	7.8	8.1	8.0	7.9
03	0815	9.5	4.5	4.0	7.5	7.5	8.0	8.0	7.7
04	0740	9.5	4.5	4.0	7.5	7.6	8.1	7.9	7.9
05	0745	9.0	4.5	4.0	7.5	7.5	8.0	8.0	8.0
06	0730	9.3	4.5	4.0	7.6	7.6	8.0	7.7	7.6
07	0740	9.4	4.5	4.0	7.6	7.6	8.2	7.4	7.3

Appendix B-3.2. PILOT SYSTEM TEMPERATURE PROFILES

Date	Time	Undiluted Waste to Neutralization Tank °C	Neutralization Tank °C	AS-1		AS-2		AS-2 Effluent °C
				Cell 1 °C	Cell 3 °C	Cell 1 °C	Cell 3 °C	
10/26/79	1600	--	35	31	30	25	24	24
27	0830	--	38	31	30	25	24	24
28	0945	--	35	35	33	27	26	26
29	1500	--	39	34	33	28	28	27
30	--	--	37	33	32	27	27	26
31	1000	--	37	33	32	27	27	26
11/01/79	1100	41	38	36	35.5	32	31.5	30.5
02	0920	39	37.5	34.5	34	31	31	31
03	0900	40	37	33	32	28	28	27
04	1115	44.5	31.5	31	30.5	26	25.5	25
05	0850	44	38	32	31	26	25	24
06	0840	39.5	35	31	30	26	25.5	25
07	1400	40.5	36	29.5	29	23.5	23.5	23
08	0815	41	37	29	28	23.5	23	22.5
09	1720	35	33	29	28	25	25	24
10	0650	40	38	32	31	27	26	25
11	0700	--	--	31	30	26	25	25
12	0900	42	37.5	32	32	26.5	26	25.5
13	1000	--	39	31	30	27	26.5	26
14	1000	--	38	32	31	27	26.5	26
15	0830	--	39	30	29.5	25.5	25	24.5
16	1030	--	39	32	31	27.5	27	26.5
17	1000	--	37	31.5	30.5	27	26.5	26
18	1100	--	37	32	30.5	28	26	27
19	1100	--	38	32.5	31.5	29	28.5	28
20	--	--	--	--	--	--	--	--
21	0745	35	33.5	33.5	33	29.5	29	28.5
22	0955	43	36.5	34.5	34	31	30.5	29.5
23	0815	42.5	39.5	36	35.5	30	29.5	29
24	0730	44	39	37	36.5	32	31.5	30.5
25	1100	37	35.5	32	31.5	28.5	28	27.5
26	0800	45.5	42	36	35.5	31.5	31	30
27	1145	44	40.5	33.5	33	30	29.5	29
28	1005	39	36.5	34	34	30	30	29.5
29	0815	46.5	38	33	33	29	28.5	28
30	0935	39.5	36.5	33	33	27	26.5	26
12/01/79	0955	35	34	30	30	26	25	25
02	0930	41	37	31	30.5	25.5	25.5	25
03	1100	46	31.5	30.5	30.5	24.5	24.5	24
04	1100	47.5	39	32.5	31.5	24	24	24
05	1000	25.5	33	33	32.5	25	25	24.5
06	1030	45	39	34	33.5	27	27	26.5
07	1000	42.5	37	33	32.5	27	27	26.5
08	1100	38	37	33	32.5	26	25.5	25.5
09	1045	42	36	32.5	31	25	25	24.5
10	1130	45	36	32	30	24	24	24
11	1000	40	34	31	30	24	23.5	23.5
12	0945	39	33	30	29	23	22.5	22.5

continued

Appendix B-3.2. (continued)

Date	Time	Undiluted Waste to Neutralization Tank °C	Neutralization Tank °C	AS-1		AS-2		AS-2 Effluent °C
				Cell 1 °C	Cell 3 °C	Cell 1 °C	Cell 3 °C	
13	0930	38	32	28	27	22.5	22	21.5
14	1030	37	33	27	26	22	21.5	21.5
15	1015	35	30	26	25	21.5	21	21
16	0940	37	29	26	25	21	21	20.5
17	0945	34	30	26	25	21	20	20
18	--	31	32	26	25	20.5	20	20
19	1500	39	36	27	26	22	21.5	21
20	0825	38	33	29	28.5	23.5	23.5	23
21	0815	38-42.5	35.5	29.5	29	23.5	23	23
22	0700	42	37	31	30.5	25.5	25	24.5
23	0850	46	40	32.5	32	27	27	26
24	0715	42	38.5	34.5	34	28.5	28	28
25	0630	41.5	37	33.5	33.5	28.5	28.5	28
26	0650	41	36.5	32	31.5	27	27	26.5
27	0640	40	35.5	30.5	30	26	25.5	25.5
28	0655	43	37.5	31.5	30.5	26	25.5	25
29	0820	42	36.5	31	30.5	26	25.5	25.5
30	0750	42.5	37	31	30.5	26	25.5	25
31	0715	42	37.5	32	31.5	26.5	26	25.5
1/01/80	0815	45.5	40	33	32	27	26.5	26
02	0650	45	37.5	32.5	32	28	27.5	27.5
03	0645	40.5	38	32.5	32	28.5	28	28
04	0700	40	37.5	30	30	29	28.5	27.5
05	0710	42.5	37.5	32	31	31	30	29
06	0710	50	41.5	32	31	30.5	29.5	28.5
07	0600	46	40	34	32.5	32	31	29.5
08	0720	38	36	31.5	31	32	30.5	29
09	0810	45	37.5	30.5	29.5	31	29.5	28.5
10	0745	46	39	31	29.5	31	29.5	28.5
11	0740	35	36	32	31	32.5	31.5	30.5
12	0800	44	39	33	32	32	30.5	29.5
13	0730	43	38	32.5	32	32	30.5	29
14	0730	45	37	33	32.5	32	31	30
15	0700	42	36	33	33	32.5	31.5	30
16	0845	37	36	34.5	34	33	31	29.5
17	0800	38	35	34	33	33	32	30.5
18	0700	35	36	32	31	32	31	30
19	0900	37	33	30	30	31	30	29
20	0900	42	34	32	31	30	29	28
21	0840	33	36	32	31	31	30	29
22	1000	32	35	32	32	32	31	30
23	1015	40	36	32	31	31	30	29
24	1515	31	30	28	27	30	29	28
25	1030	40	29	27	27	30	29	28
26	0930	36	28	28	28	31	29	28
27	0915	42	34	28	27	30	29	28
28	1015	43	34	31	30	31	29	28
29	0840	46	37	34	33	31	29.5	28.5

continued

Appendix B-3.2. (continued)

Date	Time	Undiluted Waste to Neutralization Tank °C	Neutralization Tank °C	AS-1 Cell 1 °C	AS-1 Cell 3 °C	AS-2 Cell 1 °C	AS-2 Cell 3 °C	AS-2 Effluent °C
30	0830	37.5	34	32	31.5	31	29.5	29
31	0810	34.5	35	33	32	32	30	29
2/01/80	0900	45	39	34	33	34	33	32.5
02	0950	41	37.5	34	33.5	34	33	33
03	0800	40	38	34.5	34.5	34	32.5	32.5
04	0710	41	37	35	35	34	33	33
05	0715	45	38.5	35	35	34	33	33.5
06	0710	38	35	35	35.5	35	34.5	33.5
07	0705	40	36	34.5	34.5	35	34.5	34

Appendix B-3.3 PILOT STUDY DISSOLVED OXYGEN DATA

	AS-1			AS-2		
Date	Cell 1 mg/l	Cell 2 mg/l	Cell 3 mg/l	Cell 1 mg/l	Cell 2 mg/l	Cell 3 mg/l
10/20/79	0.4	0.8	2.7			
21	1.2	2.8	3.0			
22	3.0	3.5	4.7	6.2	6.4	6.4
23						
24	0.7	1.1	3.0	6.5	6.8	7.2
25	1.1	3.1	0.3	3.4	5.3	4.8
26	1.1	4.5	1.9	7.0	7.2	7.2
27	1.1	1.2	2.3	6.8	7.2	7.5
28	0.8	2.2	4.2	6.4	4.2	0.3
29	2.2	3.0	4.0	5.0	5.0	1.8
30	3.5	2.5	2.2	5.8	6.5	4.2
31	2.0	2.1	2.2	4.0	4.2	3.2
11/01/79	0.15	0.2	0.4	2.0	2.0	2.6
02	0.7	3.4	5	5.3	2.6	3.2
03	1.8	1.6	5	6	2	5
04	0.25	0.4	0.5	3.2	2.4	7
05	1.2	0.4	1.2	0.4	1.2	5
06	0.3	0.2	0.5	0.3	1.6	4
07	0.3	0.3	1.0-1.8	3.4	4.3	4.6
08	0.3	0.3	4.0	3.8	4.6	4.8
09	0.5	0.2	2.6	2.5	4	6.4
10	2.0	0.1	1.1	0.3	0.3	2.4
11	0.25	0.25	2.0	0.3	0.8	1.8
12	0.2	0.2	2.0	0.35	0.35	0.3
13	0.8	1.0	1.1	2.4	2.6	2.8
14	0.9	1.0	1.2	2.5	3.0	3.0
15	1.5	1.8	2.0	2.5	1.8	1.6
16	1.3	1.3	1.5	1.9	2.1	4.0
17	1.4	1.2	1.6	1.6	2.1	4.1
18	1.4	1.3	1.6	1.9	2.2	4.5
19	2.2	0.9	1.8	1.8	4.5	5.2
20	1.0	1.1	1.4	1.8	0.9	1.5
21	0.2	1.1	3.5	0.7	4.5	5
22	0.2	0.25	0.3	0.15	0.2	0.9
23	0.3	0.2	0.5	6.4	6	6
24	0.2	0.2	0.2	4.0	5	6
25	0.5	0.5	3.5	0.25	0.4	0.6
26	0.3	0.2	1.8	<0.2	0.6	3
27	0.2	0.3	3.5	0.2	1.4	3.5
28	0.3	0.6	1.6	0.25	0.2	0.3
29	0.2	0.1	1.2	1.5	3	3

(continued)

Appendix B-3.3. (continued)

Date	AS-1 Cell 1 mg/1	AS-1 Cell 2 mg/1	AS-1 Cell 3 mg/1	AS-2 Cell 1 mg/1	AS-2 Cell 2 mg/1	AS-2 Cell 3 mg/1
11/30/79	0.2	0.4	0.4	3.8	5.5	5.8
12/01/79	0.2	0.8	0.85	<0.1	5.6	7.0
02	0.3	0.6	0.3	2.3	6.5	7.2
03	3.8	5.0	2.8	7.0	8.4	8.8
04	3.8	2.5	1.6	3.0	7.3	7.2
05	1.6	1.6	0.9	3.8	6.0	5.8
06	0.8	0.9	1.0	3.8	5.2	6.0
07	1.2	1.0	0.8	3.8	6.0	6.0
08	1.0	0.6	0.8	3	5	6
09	0.3	0.4	0.6	3.2	5.2	6.2
10	0.8	2.2	5.5	1.2	5.2	5.2
11	0.8	1.0	3.5	0.8	5.8	6.8
12	1.2	3.0	4.8	3.0	5.4	5.8
13	1.5	2.6	4.6	2.8	5.3	5.6
14	1.5	4.2	5.8	2.5	6.7	6.8
15	2.4	3.8	4.2	3.2	3.2	4.5
16	2.0	4.2	6.0	5.0	7.0	7.2
17	1.8	5.0	3.3	2.0	6.4	6.6
18	1.8	5.0	3.3	1.4	5.8	5.4
19	1.4	0.7	4.2	3.4	5.4	4.2
20	0.2	0.25	4.5	5	5.5	3.5
21	0.1	0.5	4.8	4.6	5.4	4.0
22	0.2	0.7	4.5	3.8	5.0	5.9
23	0.2	0.9	4	3.0	4	5
24	0.4	0.5	1.5	2.8	4	5
25	0.25	1.8	5.2	3	4	5.5
26	0.3	3.5	6	3.5	5	5.5
27	0.3	3.5	6	3.5	5	5.5
28	0.2	4.5	5.0	3.0	4.5	5
29	0.2	3.2	3.6	4.5	5.0	4.6
30	0.3	4.4	2.0	5.8	5.5	5.2
31	0.45	4.0	3.2	6.2	6.0	5.5
1/01/80	0.45	1.1	2.7	6	5	4.5
02	0.4/0.5	3/2.5	0.7/1.6	5.5	4.5	5
03	1.0	4.5	4.7	5.0	3.5	4.4
04	0.45	5.5	5.5	3.2	1.2/4.5	1.3/4.5
05	0.3	4.4	5	3.0	5.2	6.0
06	0.3	4.5	5.4	2.8	4.5	5.5
07	2.5	3.5	6.2	1.7	4.5	6.0
08	2.2	1.9	6.5	4.2	5.6	6.0
09	2.5	4.0	5.4	5.4	6.0	6.3
10	2.4	4.3	6.6	2.5	5.0	6.0

(continued)

Appendix B-3.3. (continued)

	AS-1			AS-2		
Date	Cell 1 mg/l	Cell 2 mg/l	Cell 3 mg/l	Cell 1 mg/l	Cell 2 mg/l	Cell 3 mg/l
11	1.8	3.0	5.0	2.1	4.4	5.2
12	1.2	3.2	6.4	1.7	4.1	4.7
13	3.0	4.0	4.2	2.0	4.2	4.5
14	1.2	3.2	3.5	3.2	5.0	5.6
15	1.5	1.5	3.0	1.8	4.0	4.4
16	4.0	2.4	3.8	4.0	6.6	6.0
17	1.8	4.0	5.8	3.1	5.2	5.6
18	3.2	6.0	6.0	3.5	5.4	6.3
19	4.0	5.4	6.0	4.2	6.0	7.2
20	4.2	3.6	5.0	2.0	7.2	7.2
21	4.2	3.8	5.4	3.0	5.6	6.6
22	4.2	3.8	5.4	3.0	5.6	6.6
23	2.4	4.5	5.4	2.2	7.2	5.8
24	2.4	3.6	3.0	2.8	3.4	6.0
25	1.2	9.8	10.6	0.95	1.4	7.0
26	4.6	6.0	5.8	1.2	10.1	11.0
27	2.0	2.8	4.0	2.2	4.2	6.8
28	2.0	5.8	4.8	1.0	5.6	6.0
29	2.0	0.5	5.6	3.0	4.3	2.5
30	3.0	2.7	5.0	3.0	6.0	6.3
31	2.3	0.7	4.0	3.0	5.0	5.5
2/01/80	1.0	2.6	4.0	1.0	3.6	5.0
02	2.5	0.6	3.4	2.6	4.2	5.4
03	1.5	0.7	3.0	2.1	4.7	5.3
04	1.5	0.2	0.2	1.5	4.8	4.8
05 (A.M.)	1.5	0.2	0.5	2.0	4.0	4.2
05 (P.M.)	0.8	0.2	2.5	2.0	4.0	4.2
06	2.0	0.7	2.0	1.9	3.2	4.7
07	1.8	0.8	0.8	2.1	3.0	4.1

Appendix B-3.4. BIOLOGICAL TEST SITE - BOD REMOVAL SUMMARY - AS-I

Date	Feed BOD mg/l	AS-I eff. BOD mg/l	Percent removal
10/21/79	1728	270	84.0
10/25/79	1776	344	81.0
10/30/79	1587	333	79.0
10/31/79	1659	246	85.0
11/2/79	1695	136	92.0
11/12/79	1821	267	85.0
11/14/79	2238	906	60.0
11/16/79	2100	807	62.0
11/19/79	1923	341	82.0
11/21/79	2112	269	87.0
11/23/79	1941	255	87.0
11/26/79	1896	436	77.0
11/28/79	2280	321	86.0
11/30/79	2070	245	88.0
12/3/79	2040	300	85.0
12/5/79	2394	6.9	74.0
12/7/79	2550	786	69.0
12/10/79	2080	429	79.0
12/14/79	2034	131	94.0
12/17/79	1491	145	90.0
12/19/79	1629	147	91.0
12/21/79	1680	166	90.0
12/24/79	1899	253	87.0
12/26/79	1795	134	92.0
12/28/79	1870	165	91.0
12/31/79	1780	51.0	97.0
1/2/80	2118	31.8	98.0
1/4/80	1767	40.7	98.0
1/7/80	1812	96.6	95.0
1/9/80	1968	48.3	98.0
1/11/80	1722	35	98.0
1/14/80	1530	60	96.0
1/16/80	1614	105	93.0
1/18/80	1371	38.4	97.0
1/21/80	1290	12.6	99.0
1/23/80	1398	61.0	96.0
1/25/80	1503	33.9	98.0
1/28/80	1614	177	89.0
1/30/80	1758	90	95.0
2/1/80	1560	63	97.0

(continued)

Appendix B-3.4. (continued)

Date	Feed BOD mg/l	AS-I eff. BOD mg/l	Percent removal
2/2/80	1458	66	95.0
2/4/80	1590	81	95.0
2/5/80	1443	151	90.0
2/6/80	1659	190	89.0
2/7/80	1623	155	90.0

Appendix B-3.5. BIOLOGICAL TREATMENT SYSTEM - PHENOL REMOVAL ACROSS AS-I

Date	Phenol concentration, mg/l		
	Diluted feed	AS-I effluent	% Removal
10/25/79	880	1.5	99.8
10/26/80	550	1.2	99.8
10/30/79	811	12.2	98.5
10/31/79	805	2.7	99.7
11/1/79	820	26.4	96.8
11/2/79	800	37.6	95.3
11/5/79	800	3.05	99.6
11/7/79	694	28	96.0
11/12/79	890	4.14	99.5
11/16/79	920	2.9	99.7
11/19/79	780	2.3	99.7
11/21/79	848	5.0	99.4
11/23/79	857	3.67	99.6
11/26/79	573	2.81	99.5
11/28/79	604	0.13	99.9
12/3/79	720	7.4	99.0
12/5/79	800	70	91.3
12/7/79	870	90	89.7
12/10/79	640	16	97.5
12/24/79	800	5.6	99.3
12/17/79	550	3.8	99.3
12/21/79	610	5.7	99.1
12/24/79	580	0.8	99.9
12/26/79	510	0.4	99.9
12/28/79	560	2.6	99.5
12/31/79	530	1.4	99.7
1/2/80	638	<.01	100.0
1/7/80	590	1.2	99.8
1/11/80	650	0.4	99.9
1/14/80	440	0.6	99.9
1/18/80	540	0.4	99.9
1/21/80	520	2.1	99.6
1/25/80	490	0.5	99.9
1/28/80	540	0.2	99.9
1/30/80	540	1.2	99.8
2/1/80	560	1.4	99.8
2/2/80	440	<.2	99.9
2/3/80	560	0.8	99.9
2/4/80	580	0.4	99.9
2/5/80	580	1.2	99.8
2/6/80	600	1.4	99.8
2/7/80	540	0.8	99.9

Appendix B-3.6. AS-I PHOSPHORUS AND THIOCYANATE DATA

Date	Effluent total P as P, mg/l	Effluent ortho P as P[1], mg/l	Influent SCN, mg/l	Effluent SCN, mg/l
10/25/79	6.0	-	480	370
10/30/79	2.6	-	350	360
10/31/79	2.5	-	330	350
11/1/79	4.5	-	330	360
11/5/79	10.8	4.1	370	350
11/16/79	5.3	-	360	380
11/19/79	3.2	-	320	350
12/3/79	4.2	4.0	455	425
12/5/79	5.7	4.5	440	480
12/7/79	6.3	3.5	-	-
12/10/79	1.5	-	420	445
12/12/79	4.7	1.9	-	-
12/14/79	1.0	0.5	385	440
12/17/79	1.7	0.4	390	425
12/19/79	5.8	2.7	495	440
12/21/79	9.0	5.6	435	430
12/24/79	-	-	525	470
12/26/79	10.1	9.4	545	515
12/28/79	-	-	490	480
12/31/79	-	-	515	3.0
1/2/80	9.0	8.1	570	4.6
1/4/80	9.4	6.6	600	23
1/7/80	9.2	5.8	520	390
1/9/80	8.8	7.1	610	515
1/11/80	8.6	6.6	535	525
1/14/80	10.2	7.6	590	410
1/16/80	11.2	7.8	580	130
1/18/80	11.3	7.7	550	11
1/21/80	9.4	8.2	580	8
1/23/80	8.4	7.5	550	4
1/25/80	8.7	7.7	500	2
1/28/80	5.6	5.8	470	3
1/30/80	7.8	7.7	550	480
2/1/80	6.6	-	470	525
2/2/80	7.1	8.3	475	470
2/3/80	8.1	-	445	345
2/4/80	7.7	6.8	490	100
2/5/80	7.1	6.8	490	71
2/6/80	7.8	7.7	455	235
2/7/80	8.3	-	425	330

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[1] Values based on field measurements.

Appendix B-3.7. AS-1 FIELD SOLIDS DATA

Date	MLSS mg/l	MLVSS mg/l	SVI	Setting[a] rate ft/min	Return sludge total solids mg/l	Return sludge total volatile solids mg/l
10/21/79	2,000	-	-	-	-	-
22	-	-	-	-	-	-
23	-	-	-	-	-	-
24	1,983	-	131	0.19	12,100	-
25	2,500	-	104	0.19	13,400	-
26	2,050	-	129	0.19	11,800	-
27	2,277	-	121	0.22	12,400	-
28	2,370	-	118	0.19	13,900	-
29	2,370	-	122	0.18	12,400	-
30	2,520	-	125	0.16	11,000	-
31	2,780	-	111	0.17	-	-
11/01/79	2,500	-	124	0.16	13,600	-
02	2,600	-	-	-	15,000	-
03	2,900	-	107	0.10	-	-
04	3,380	-	112	0.11	12,800	-
05	3,050	-	105	0.17	-	-
06	3,015	2,620	-	-	20,900	14,600
07	2,640	2,630	117	-	21,000	14,600
08	3,480	3,230	-	-	12,500	6,500
09	2,500	2,260	-	-	11,900	-
10	2,420	2,110	101	0.16	12,900	7,100
11	2,500	2,100	125	0.15	11,300	6,100
12	2,350	2,000	-	-	11,600	6,200
13	2,550	2,180	125	0.10	11,200	7,000
14	2,670	2,300	101	0.08	11,000	6,600
15	2,420	2,070	125	0.01	12,000	7,500
16	3,000	2,670	137	0.06	12,500	7,500
17	2,833	2,500	121	0.06	14,100	9,000
18	3,170	2,870	112	0.06	13,600	8,300
19	3,167	2,867	-	-	12,400	7,300
20	3,220	2,920	-	-	9,940	4,750
21	3,530	3,160	121	0.13	12,700	7,420
22	3,280	3,280	155	0.03	12,600	7,160
23	2,590	2,500	58	0.07	11,400	5,900
24	1,590	1,590	81	0.04	11,200	5,460
25	2,050	1,760	48	0.10	-	-
26	1,900	1,870	83	0.06	12,200	6,200
27	2,000	-	55	0.07	12,800	6,700

[a] ft/min x 0.3048 = m/min

(continued)

Appendix B-3.7. (continued)

Date	MLSS mg/l	MLVSS mg/l	SVI	Settling[a] rate ft/min	Return sludge total solids mg/l	Return sludge total volatile solids mg/l
28	1,970	1,830	-	-	12,000	6,100
29	1,680	930	65	0.06	10,800	4,640
30	2,570	2,200	50	-	11,100	5,070
12/01/79	2,420	2,300	49	.07	12,920	6,930
02	2,700	1,767	77	.05	16,900	10,700
03	3,785	2,400	25	-	13,960	7,600
04	1,267	1,100	24	-	13,900	8,060
05	2,005	1,510	39	-	24,400	17,500
06	2,350	1,740	26	-	14,670	8,290
07	1,930	1,675	-	-	14,720	7,330
08	1,690	1,650	48	-	15,880	8,310
09	1,980	1,850	45	-	20,200	10,100
10	3,200	2,810	34	-	16,800	9,460
11	3,333	3,000	-	-	12,900	5,900
12	3,500	3,300	54	0.07	14,050	7,270
13	3,530	3,300	56	0.07	15,600	8,800
14	3,820	3,670	51	0.1	14,700	8,090
15	3,680	3,520	50	0.1	13,200	6,650
16	3,280	3,150	53	0.04	13,300	6,750
17	2,730	2,620	57	0.06	13,500	6,900
18	2,570	2,330	49	0.06	12,000	5,600
19	2,400	2,230	49	0.05	11,900	5,500
20	2,550	2,280	38	-	11,800	5,400
21	2,567	2,470	45	-	12,040	5,780
22	2,780	2,730	55	0.03	13,000	3,760
23	2,450	2,230	47	0.03	12,300	5,790
24	2,450	2,300	18	-	12,200	5,495
25	2,600	2,580	44	-	13,000	6,170
26	2,730	2,570	90	0.13	13,100	6,270
27	2,800	2,800	93	0.07	14,300	7,250
28	2,875	2,875	129	0.04	14,100	7,120
29	3,470	2,900	154	0.03	12,900	5,900
30	3,400	3,120	-	-	13,200	6,180
31	3,380	3,220	93	-	13,900	6,545
1/01/80	3,420	3,180	96	0.07	14,100	6,815
02	3,250	3,025	95	0.10	14,200	6,880
03	3,200	3,100	-	-	14,200	6,860
04	3,120	2,950	104	0.13	13,000	5,737

[a] ft/min x 0.3048 = m/min

(continued)

Appendix B-3.7. (continued)

Date	MLSS mg/l	MLVSS mg/l	SVI	Settling[a] rate ft/min	Return sludge total solids mg/l	Return sludge total volatile solids mg/l
05	3,300	3,100	96	0.09	14,600	7,670
06	3,220	3,180	100	0.09	14,400	7,360
07	3,275	3,025	100	0.10	15,800	8,740
08	3,450	3,400	120	0.10	15,900	8,800
09	3,650	3,575	105	0.10	14,400	7,210
10	4,050	4,000	117	0.10	14,700	7,310
11	4,300	4,125	92	0.11	15,700	8,040
12	4,475	4,300	97	0.11	16,000	8,230
13	4,725	4,400	97	0.10	16,700	8,690
14	4,500	4,325	100	0.10	15,800	8,290
15	4,625	4,050	100	0.10	15,700	8,530
16	4,650	4,580	89	0.09	16,700	9,740
17	5,100	4,450	76	0.09	18,330	8,750
18	4,400	3,700	95	0.12	14,680	6,655
19	4,450	4,100	80	0.12	14,660	8,070
20	4,825	3,975	75	0.13	16,295	9,460
21	4,400	3,975	84	0.12	15,965	8,575
22	4,750	4,275	80	0.12	16,200	8,550
23	5,250	4,750	72	0.12	16,560	8,735
24	-	-	70	0.16	-	-
25	5,950	5,700	62	0.09	15,710	7,645
26	5,750	5,125	83	0.14	17,861	9,850
27	6,300	5,975	71	0.09	18,935	10,190
28	7,200	6,525	76	0.06	20,755	11,945
29	6,300	5,800	101	0.03	23,287	14,815
30	6,650	6,250	84	0.07	21,700	12,470
31	6,275	6,000	121	0.01	20,877	11,737
2/01/80	6,250	6,025	120	0.01	21,180	12,464
02	5,825	5,525	80	0.07	20,675	12,336
03	5,975	5,750	109	0.01	20,384	12,214
04	5,950	5,800	129	0.01	20,949	12,681
05	6,175	5,875	132	0.01	20,353	12,354
06	6,550	6,225	-	-	21,603	13,724
07	6,700	6,400	88	0.05	19,213	11,698

[a] ft/min x 0.3048 = m/min

Appendix B-3.8. SELECTED TEST PERIODS OF AS-I OPERATION - FOLLANSBEE

Date	SS mg/l	VSS mg/l	NH_3-N as N mg/l	TKN as N mg/l	NO_2/NO_3-N mg/l	Phenol mg/l	CN_T mg/l	CN_A mg/l	O & G mg/l	TOC mg/l	Total P as P mg/l	$S^=$ mg/l	BOD mg/l	SCN mg/l	Color APHA Units
System Feed - Period 1															
11/12/79	53	15	1,420	1,588		890	10.8	1.0		1,110		32		384	
13										1,180					
14	110	42	1,760	2,008			10.0	<0.05		1,250			2,238	410	1,800
15										1,050					
16	41	30	1,367	1,423		920	9.6	0.2	4	1,250	7.3	34	2,100	360	
17										1,410					
18										1,000					
19	37	32	692	793		780	6.5	0.7	5	1,040	6.2	32	1,923	320	600/700
AS-1 Effluent - Period 1															
11/12/79	77	48	1,086	1,249		4.1	7.2	0.2		320		2.4		370	
13										260					
14	189	156	1,574	1,605			10.0	0.5		440			906	400	1,500
15										330					
16	198	158	1,364	1,605		2.9	8.6	0.3	5	510	5.3	39	807	380	
17										450					
18										630					
19	86	77	955	1,025		2.3	5.7	<0.05	5	305	3.2	3.5	341	350	600
System Feed - Period 2															
12/19/79	51	24	616	661			0.5	<0.05		1,400	11.4		1,629	495	400/500
20										1,230					
21	64	16	420	501		610	0.9	<0.05	16	1,290	11.1	24	1,680	435	300/400
22										1,200					
23										1,370					
24	70	20	319	482		580	0.8	<0.05	105	1,390			1,899	525	400/500
25										1,250					
26	58	19	280	372		510	0.6	<0.05		1,380	6.3		1,795	545	

continued

Appendix B-3.8. (continued)

Date	SS mg/l	VSS mg/l	NH_3-N as N mg/l	TKN as N mg/l	NO_2/NO_3-N mg/l	Phenol mg/l	CN_T mg/l	CN_A mg/l	O & G mg/l	TOC mg/l	Total P as P mg/l	$S^=$ mg/l	BOD mg/l	SCN mg/l	Color APHA Units
AS-1 Effluent - Period 2															
12/19/79	98	65	605	622	<1		1.8	1.4		175	5.8		147	440	800/1,000
20										190					
21	135	81	437	661		5.7	0.2	<0.05	4	155	9.0	4.1	166	430	800/1,000
22										205					
23										215					
24	150	100	353	518		0.8	0.7	<0.05	5	235			253	470	1,000/1,200
25										250					
26	102	52	358	431		0.4	0.3	<0.05		200	10.1		134	515	
System Feed - Period 3															
1/23/80	64	19	280	350			14.4	<0.05		940	10.0		1,398	550	
24										910					
25	74	15	220	340		490	15.2	<0.05	8	950	10.2	28.2	1,503	500	180/210
26										1,020					
27										1,040					
28	93	27	504	616		540	21.2	<0.05	16	1,050	9.8		1,614	470	
AS-1 Effluent - Period 3															
1/23/80	73	27	400	420	1.6		22.0	<0.05		176	8.4		61	4	
24										172					
25	87	31	330	360		0.5	20.2	<0.05	9	135	8.7		34	2	600/800
26										112					
27										127					
28	103	44	420	479		0.2	24.0	<0.05	12	135	5.6		177	3	

continued

Appendix B-3.8 (continued)

Date	SS mg/l	VSS mg/l	NH_3-N as N mg/l	TKN as N mg/l	NO_2/NO_3-N mg/l	Phenol mg/l	CN_T mg/l	CN_A mg/l	O & G mg/l	TOC mg/l	Total P as P mg/l	$S^=$ mg/l	BOD mg/l	SCN mg/l	Color APHA Units
System Feed - Period 4															
1/30/80	67	14	540	605		540	27.9	<0.05		1,240	9.3		1,758	550	
31										1,160					
2/01/80	77	19	636	697		560	32.4	5.76	7	1,197	7.6	28	1,560	470	250/300
02	88	29	652	728		440	20.5	<0.05	7	1,125	8.6	28	1,458	475	200/250
03	80	20	291	389		560	4.6	<0.05	4	1,116	8.0	28	1,500	445	150/200
04	205	80	297	375		580	4.2	<0.05	15	1,170	8.0	30	1,590	490	150/200
05	160	70	316	409		580	6.0	1.7	24	1,197	8.0	33	1,443	490	200/300
06	30	30	288	330		600	6.3	2.4	15	1,200	10.5	34	1,659	455	300/400
07	20	19	283	364		540	8.5	<0.05	18	1,203	10.6	23	1,623	425	200/300
AS-1 Effluent - Period 4															
1/30/80	207	111	700	767		1.2	17.4	<0.05		260	7.8		90	480	
31										296					
2/01/80	206	90	619	694	1.2	1.4	13.5	5.7	6	360	6.6	5.2	43	525	1,200/1,400
02	198	85	647	724	1.0	<0.2	15.2	1.4	4	140	7.1	3.6	66	470	1,250/1,500
03	220	106	596	661		0.8	5.6	<0.05	10	250	8.1	6.4	75	345	1,000/1,200
04	280	105	521	585	1.1	0.4	16.9	2.9	10	182	7.7	2.2	81	100	1,000/1,200
05	260	140	437	487	1.5	1.2	14.7	<0.05	20	230	7.1	1.4	151	71	1,000/1,200
06	120	100	381	386	1.3	1.4	8.6	0.57	9	243	7.8		190	235	500/600
07	70	65	336	301	1.5	0.8	7.4	<0.05	12	266	8.3		155	330	1,200/1,400

Appendix B-3.9. PILOT PLANT DATA SUMMARY ON BOD REMOVAL, MLVSS, SVI, SETTLING RATE, RETENTION TIME AND F/M RATIO - AS-I

Date	Influent BOD, mg/l	BOD percent removal	MLVSS, mg/l	SVI	Settling rate, ft/min	Hydraulic retention time, hours	F/M day^{-1}
10/21/79	1,728	84.0	-	-	-	26.1	-
10/25	1,776	81.0	-	104	0.19	26.1	-
10/30	1,587	79.0	-	125	0.16	33.0	-
11/2	1,695	92.0	-	-	-	30.0	-
11/12	1,821	85.0	2,000	-	-	24.8	0.881
11/14	2,238	60.0	2,300	101	0.08	33.0	0.708
11/16	2,100	62.0	2,670	137	0.06	26.1	0.723
11/19	1,923	82.0	2,867	-	-	26.1	0.617
11/21	2,112	87.0	3,160	121	0.13	41.3	0.388
11/23	1,941	87.0	2,500	58	0.07	25.8	0.722
11/26	1,896	77.0	1,870	83	0.06	25.5	0.954
11/28	2,280	86.0	1,830	-	-	27.5	1.087
11/30	2,070	88.0	2,200	50	-	25.8	0.875
12/3	2,040	85.0	2,400	25	-	25.5	0.800
12/5	2,394	74.0	1,510	39	-	25.5	1.492
12/7	2,550	69.0	1,675	-	-	26.6	1.374
12/10	2,080	79.0	2,810	34	-	33.0	0.538
12/14	2,034	94.0	3,670	51	0.10	44.0	0.302
12/17	1,491	90.0	2,620	57	0.06	43.4	0.315
12/19	1,629	91.0	2,230	49	0.05	44.0	0.398
12/21	1,680	90.0	2,470	45	-	40.6	0.402
12/24	1,899	87.0	2,300	47	0.03	42.7	0.464
12/26	1,795	92.0	2,570	90	0.13	43.1	0.389
12/28	1,870	91.0	2,875	129	0.04	40.6	0.384
12/31	1,780	97.0	3,220	93	-	45.8	0.290
1/2/80	2,118	98.0	3,025	95	0.10	47.6	0.353
1/4	1,767	98.0	2,950	104	0.13	53.8	0.267

ft/min x 0.3048 = m/min

continued

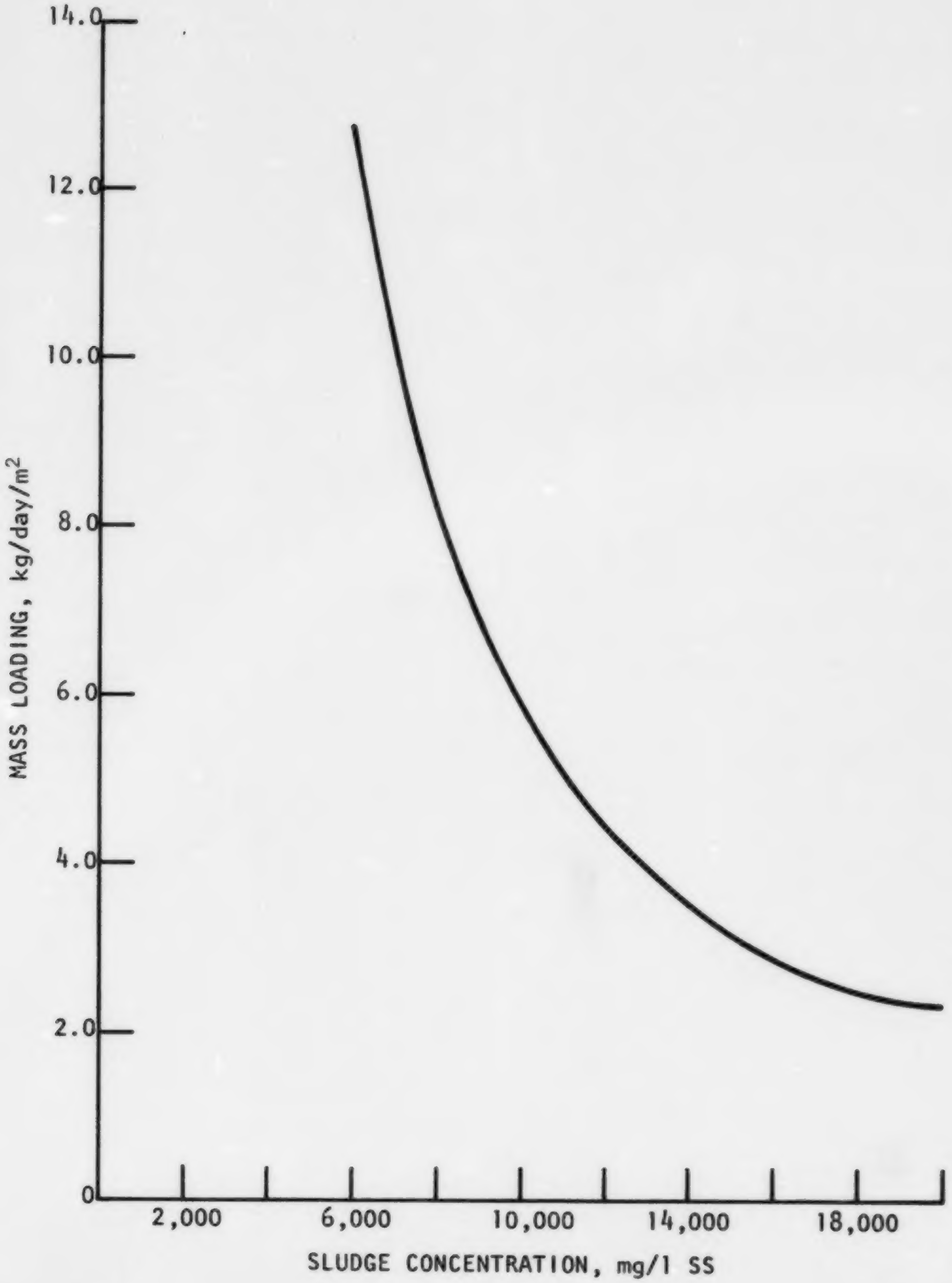
Appendix B-3.9 (continued)

Date	Influent BOD, mg/l	BOD percent removal	MLVSS, mg/l	SVI	Settling rate, ft/min	Hydraulic retention time, hours	F/M day^{-1}
1/7/80	1,812	95.0	3,025	100	0.10	40.3	0.357
1/9	1,968	98.0	3,575	105	0.10	43.8	0.302
1/11	1,722	98.0	4,125	92	0.11	33.2	0.302
1/14	1,530	96.0	4,325	100	0.10	33.9	0.250
1/16	1,614	93.0	4,580	89	0.09	30.0	0.282
1/18	1,371	97.0	3,700	95	0.12	32.6	0.273
1/21	1,290	99.0	3,975	84	0.12	33.5	0.232
1/23	1,398	96.0	4,750	72	0.12	35.6	0.198
1/25	1,503	98.0	5,700	62	0.09	35.4	0.179
1/28	1,614	89.0	6,525	76	0.06	27.7	0.214
1/30	1,758	95.0	6,250	84	0.07	26.2	0.258
2/1	1,560	97.0	6,025	120	0.01	27.1	0.229
2/2	1,458	95.0	5,525	80	0.07	25.8	0.245
2/4	1,590	95.0	5,800	129	0.01	26.8	0.245
2/5	1,443	90.0	5,875	132	0.01	25.5	0.231
2/6	1,659	89.0	6,225	-	-	25.4	0.252
2/7	1,623	90.0	6,400	88	0.05	26.3	0.231

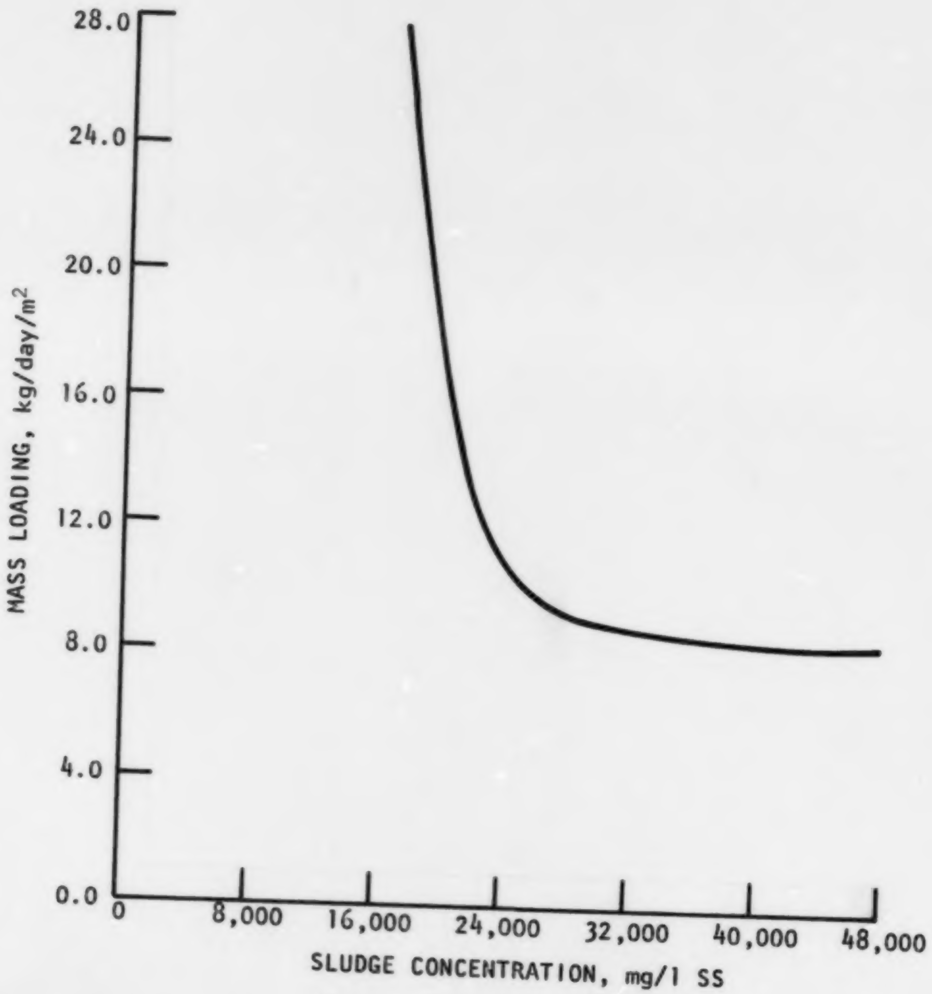
ft/min x 0.3048 = m/min

Appendix B-3.10. OXYGEN UPTAKE TEST RESULTS FOR EACH CELL IN AS-1 AND AS-2

Date	Time	Location	MLVSS (mg/l)	Temp. (°C)	O_2 Demand (mg/l/hr)	Mixed Liquor Flow (gpm)	Detention Time (hrs)	F/M Ratio
2/3/80	1300	AS-1, Cell 1	5,750	31.2	306	1.99	4.4	0.24# BOD/day/# MLVSS
2/3/80	1330	AS-1, Cell 2	5,750	32.0	103	1.99	8.8	
2/3/80	1400	AS-1, Cell 3	5,750	32.0	42	1.99	13.2	
2/6/80	0900	AS-1, Cell 1	6,225	32.0	381	2.02	4.3	0.24# BOD/day/# MLVSS
2/6/80	0930	AS-1, Cell 2	6,225	32.0	304	2.02	8.7	
2/6/80	1000	AS-1, Cell 3	6,225	31.2	168	2.02	13.0	
2/5/80	0900	AS-2, Cell 1	6,200	29.7	56	1.91	4.3	0.024# NH_3-N/day/#MLVSS
2/5/80	0930	AS-2, Cell 2	6,200	28.7	21	1.91	8.6	
2/5/80	1000	AS-2, Cell 3	6,200	29.0	13	1.91	12.8	
2/7/80	0900	AS-2, Cell 1	5,700	32.0	114	1.93	4.2	0.023# NH_3-N/day/#MLVSS
2/7/80	0930	AS-2, Cell 2	5,700	30.5	111	1.93	8.5	
2/7/80	1000	AS-2, Cell 3	5,700	29.7	19	1.93	12.7	



Appendix B-3.11 Flux concentration curve for RS-1 on 2/3/80.



Appendix 3.12 Flux concentration curve for RS-2 on 2/6/80.

Appendix B-3.13. PILOT PLANT DATA FROM FOUR SELECTED OPERATING PERIODS (STAGE I)

Date	Influent BOD, mg/l	Effluent BOD, mg/l	MLVSS mg/l	Detention time, hours	Flow ℓ/min.	F/M (day^{-1})	SVI	Settling rate (ft/min)	Effluent SS, mg/l
1/30	1,758	90	6,250	26	3.80	0.25	84	0.07	136
2/1	1,560	43	6,025	28	3.52	0.22	120	0.01	79
2/2	1,458	66	5,525	26	3.80	0.24	80	0.07	112
2/3	-	-	5,750	26	3.80	-	109	0.01	96
2/4	1,590	81	5,800	27.5	3.60	0.24	129	0.01	73
2/5	1,443	151	5,875	25	3.96	0.23	132	0.01	69
2/6	1,659	190	6,225	24.7	4.00	0.25	-	-	74
2/7	1,623	155	6,400	26.6	3.72	0.23	88	0.05	70
AVERAGE	1,584	111	5,981	26.2	3.77	0.24	106	.033	89
12/19	1,629	147	2,230	44	2.25	0.40	49	0.05	35
12/20	-	-	2,280	43	2.28	-	38	-	-
12/21	1,680	166	2,430	40	2.44	0.40	45	-	39
12/22	-	-	2,730	44	2.24	-	55	0.03	-
12/23	-	-	2,230	42	2.32	-	47	0.03	-
12/24	1,899	253	2,300	42	2.32	0.46	18	-	69
12/25	-	-	2,580	43	2.28	-	44	-	-
12/26	1,795	134	2,570	42	2.30	0.39	90	0.13	34
AVERAGE	1,750	175	2,424	42.5	2.30	0.413	48.3	0.036	44
11/12	1,821	267	2,000	24.7	4.0	0.885	-125	-	77
11/13	-	-	2,180	26	3.8	-	125	0.1	-
11/14	2,238	906	2,300	33	3.0	0.708	101	.08	158
11/15	-	-	2,070	26	3.8	-	125	0.01	-
11/16	2,100	807	2,670	26	3.8	0.726	137	.06	160

(continued)

Appendix B-3.13 (continued)

Date	Influent BOD, mg/l	Effluent BOD, mg/l	MLVSS mg/l	Detention time, hours	Flow ℓ/min.	F/M (day^{-1})	SVI	Settling rate (ft/min)	Effluent SS, mg/l
11/17	-	-	2,500	26.7	3.7	-	121	0.06	-
11/18	-	-	2,870	27.5	3.6	-	112	0.06	-
11/19	1,923	341	2,867	26	3.8	0.619	120	-	73
AVERAGE	2,020	580	2,432	27	3.7	0.735	120	0.062	-
11/23	1,941	255	2,500	25.8	3.84	0.72	58	.07	42
11/24	-	-	1,590	25.5	3.88	-	81	.04	-
11/25	-	-	1,760	75	1.32	-	48	.10	-
11/26	1,896	436	1,870	25.5	3.88	0.95	83	.06	57
11/27	-	-	1,850	35.4	2.80	-	55	.07	-
11/28	2,280	321	1,830	27.5	3.60	1.08	60	-	62
AVERAGE	2,039	337	1,900	28	3.6	0.917	64.17	.067	53

Appendix B-4.1. AS-II DILUTED FEEDWATER QUALITY
JANUARY 9 - FEBRUARY 7, 1980

Date	Dilution factor	SS, mg/l	VSS, mg/l	NH_3-N, mg/l	TKN as N, mg/l	NO_2/NO_3 as N, mg/l	Phenol, mg/l	CN_T, mg/l	CN_A, mg/l	O&G, mg/l	TOC, mg/l	TP as P, mg/l	S, mg/l	BOD, mg/l	SCN, mg/l	Color APHA Units
1/9/80	0.109	11	6	68	-	<1	-	2.7	0.3	-	28	1.0	-	5	56	-
1/10	0.143	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-
1/11	0.145	10	8	70	78	-	<0.1	1.8	1.1	2	35	1.2	0.1	5	76	150/200
1/12	0.246	-	-	-	-	-	-	-	-	-	57	-	-	-	-	-
1/13	0.350	-	-	-	-	-	-	-	-	-	77	-	-	-	-	-
1/14	0.341	34	16	150	169	-	<0.1	6.1	3.1	3	68	3.5	-	20	140	-
1/15	0.344	-	-	-	-	-	-	-	-	-	62	-	-	-	-	-
1/16	0.337	47	23	206	221	-	-	2.0	<0.05	-	51	3.8	-	35	44	-
1/17	0.330	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-
1/18	0.326	34	20	228	267	<1	0.1	3.6	<0.05	3	56	3.7	0.7	13	4	150/200
1/19	0.345	-	-	-	-	-	-	-	-	-	59	-	-	-	-	-
1/20	0.341	-	-	-	-	-	-	-	-	-	52	-	-	-	-	-
1/21	0.341	26	14	171	174	<1	0.7	5.6	<0.05	2	38	3.2	-	4	3	-
1/22	0.360	-	-	-	-	-	-	-	-	-	58	-	-	-	-	-
1/23	0.400	29	19	160	168	<1	-	8.8	<0.05	-	70	3.4	-	24	2	-
1/24	0.380	-	-	-	-	-	-	-	-	-	65	-	-	-	-	-
1/25	0.472	41	24	156	170	-	0.2	9.5	<0.05	4	64	4.1	0.9	16	<1	250/300
1/26	0.531	-	-	-	-	-	-	-	-	-	59	-	-	-	-	-
1/27	0.514	-	-	-	-	-	-	-	-	-	65	-	-	-	-	-
1/28	0.556	57	30	234	266	-	0.1	13.3	<0.05	7	75	3.1	-	97	2	-
1/29	0.573	-	-	-	-	-	-	-	-	-	190	-	-	-	-	-
1/30	0.194	40	21	136	153	-	0.2	3.4	<0.05	-	50	1.5	-	17	93	-
1/31	0.281	-	-	-	-	-	-	-	-	-	83	-	-	-	-	-
2/1	0.341	70	31	211	237	<1	0.5	5.4	1.9	2	123	2.3	1.8	15	180	400/500
2/2	0.222	44	22	144	161	<1	<0.1	3.4	0.3	<1	31	1.6	0.8	15	105	200/300
2/3	0.265	58	28	158	175	-	0.2	1.5	<0.05	3	66	2.1	1.7	20	91	200/300
2/4	0.313	88	33	163	183	<1	0.1	5.3	0.9	3	57	2.4	0.7	25	31	300/400
2/5	0.367	95	51	160	179	<1	0.4	5.4	<0.05	7	84	2.6	0.5	55	26	300/400
2/6	0.440	53	44	168	170	<1	0.6	3.8	0.26	4	109	3.4	0.5	84	105	200/300
2/7	0.430	30	28	144	164	<1	0.3	3.2	<0.05	5	114	3.6	1.2	67	140	500/600
Ave.		45	25	160	183	<1	0.3	5.0	0.49	3	68	2.7	0.9	30	65	300
Range		10-95	6-51	68-234	78-267		<0.1-0.7	1.5-13.3	<0.05-3.1	<1-7	28-190	1.0-4.1	0.1-1.8	4-97	<1-180	150-600

Appendix B-4.2. AS-2 FIELD SOLIDS DATA

Date	MLSS mg/l	MLVSS mg/l	SVI	Settling[1] Rate ft/min	Return Sludge Total Solids mg/l	Return Sludge Total Volatile Solids mg/l
10/27/79	538	-	-	-	-	-
28	540	-	-	-	-	-
29	514	-	-	-	10,100	-
30	396	-	-	-	9,400	-
31	328	-	-	-	-	-
11/01/79	248	-	-	-	-	-
02	418	-	-	-	-	-
03	3,340	-	13	-	-	-
04	2,840	-	16	-	11,400	-
05	2,940	2,540	20	-	-	-
06	2,080	1,710	-	-	11,200	4,800
07	2,430	2,320	31	-	10,300	4,100
08	3,350	2,960	-	-	11,200	4,800
09	2,870	2,390	-	-	14,100	-
10	3,240	2,720	32	0.47	12,910	6,400
11	3,300	2,940	39	0.39	18,460	11,600
12	3,125	2,600	-	-	12,300	6,200
13	3,263	2,960	46	0.40	9,818	4,700
14	2,963	2,640	63	0.40	10,630	6,000
15	3,933	3,433	51	0.35	10,890	6,100
16	4,570	4,070	39	0.30	14,037	8,500
17	3,970	3,580	45	0.18	15,300	9,500
18	3,570	3,080	45	0.18	14,670	8,900
19	3,570	3,080	-	-	13,160	7,400
20	3,570	3,130	-	-	12,830	7,035
21	3,720	3,280	58	0.05	13,500	7,600
22	3,680	3,580	57	0.15	13,300	7,260
23	3,700	3,580	68	0.12	12,900	6,900
24	3,700	3,520	68	0.13	12,800	6,730
25	4,480	4,070	62	0.16	13,600	7,400
26	4,770	4,250	54	0.17	14,400	8,000
27	5,000	-	57	0.15	14,600	8,100
28	5,050	4,420	-	-	-	-
29	4,800	4,000	49	0.16	14,400	7,550
30	5,880	4,625	41	0.12	12,400	7,116
12/01/79	4,780	4,325	59	0.12	12,270	-
02	5,330	5,100	54	0.11	10,870	7,250
03	8,000	5,300	51	0.10	10,600	7,160

continued

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[1] ft/min x 0.3048 = m/min

Appendix B-4.2 (continued)

Date	MLSS mg/l	MLVSS mg/l	SVI	Settling[1] Rate ft/min	Return Sludge Total Solids mg/l	*Return Sludge Total Volatile Solids mg/l
04	4,800	4,500	59	0.10	10,300	7,110
05	5,900	5,475	57	0.10	9,030	6,200
06	5,475	4,730	54	0.13	9,680	6,730
07	5,780	5,280	59	0.12	11,300	7,980
08	4,575	4,500	60	0.11	10,700	7,330
09	5,850	5,075	78	0.09	10,820	7,440
10	5,925	5,200	56	0.09	11,800	8,260
11	5,850	5,400	-	-	9,710	6,560
12	6,325	5,780	55	0.11	10,050	6,780
13	6,400	5,800	58	0.11	10,800	7,500
14	6,300	5,850	66	0.1	8,900	5,820
15	6,580	5,920	64	0.13	10,900	7,700
16	6,250	5,720	75	0.07	10,700	7,530
17	6,400	5,450	72	0.10	10,900	7,600
18	6,250	5,750	78	0.05	10,200	7,200
19	5,920	5,320	139	0.008	10,400	7,300
20	6,300	5,700	128	0.06	10,300	7,110
21	6,150	5,520	143	0.006	10,400	7,300
22	6,800	6,320	72	0.04	11,040	7,980
23	7,170	6,280	92	0.04	11,800	8,800
24	7,000	6,370	95	0.03	11,600	8,610
25	7,050	6,620	102	0.01	10,900	7,810
26	7,050	6,400	111	0.01	11,300	8,150
27	6,900	6,450	67	0.06	11,300	7,960
28	7,050	6,680	72	0.06	11,300	7,870
29	7,400	6,150	68	0.06	11,400	7,950
30	6,900	6,150	71	0.05	11,500	8,130
31	7,050	6,330	59	0.10	11,600	8,200
1/01/80	7,020	6,250	53	0.10	11,300	7,950
02	6,750	6,120	57	0.10	11,700	8,300
03	6,950	6,200	-	-	13,000	9,450
04	7,070	6,350	58	0.10	12,300	8,930
05	7,220	6,500	59	0.10	13,700	10,500
06	7,200	6,450	59	0.08	13,200	10,400
07	7,350	6,675	59	0.11	16,200	13,300
08	7,425	6,950	59	0.10	17,200	13,900
09	7,900	7,350	56	0.11	16,700	13,100
10	8,250	7,600	54	0.08	16,400	12,700

[1] ft/min x 0.3048 = m/min

continued

Appendix B-4.2 (continued)

Date	MLSS mg/l	MLVSS mg/l	SVI	Settling[1] Rate ft/min	Return Sludge Total Solids mg/l	Return Sludge Total Volatile Solids mg/l
11	8,650	7,880	54	0.11	18,300	14,200
12	8,500	7,780	55	0.11	19,800	15,400
13	8,780	7,950	53	0.10	20,300	15,400
14	8,875	8,080	55	0.10	20,100	14,600
15	8,980	7,620	47	0.10	21,200	15,500
16	9,080	8,220	44	0.12	18,400	12,900
17	8,700	7,620	43	0.15	19,110	13,500
18	7,950	6,850	40	0.13	19,355	13,240
19	8,200	7,400	40	0.15	18,400	11,880
20	8,250	6,700	40	0.17	19,720	14,490
21	7,325	6,650	45	0.17	18,175	12,165
22	7,550	6,625	45	0.17	18,580	12,658
23	7,250	5,750	47	0.25	18,490	12,740
24	-	-	-	-	-	-
25	7,750	6,750	45	0.17	17,895	11,765
26	7,425	6,450	47	0.16	26,840	19,618
27	7,200	6,425	49	0.16	20,260	13,100
28	7,900	7,000	48	0.12	20,140	12,850
29	7,250	6,650	52	0.17	19,550	12,020
30	7,600	6,500	45	0.18	17,400	9,600
31	7,450	6,300	40	0.23	21,168	13,725
2/01/80	7,100	6,300	44	0.23	19,540	12,421
02	6,700	5,650	42	0.23	17,603	11,622
03	7,000	6,000	38	0.23	19,688	13,183
04	7,300	6,200	34	0.20	18,772	12,174
05	7,300	6,200	35	0.23	18,230	11,740
06	6,900	5,800	-	-	18,090	11,470
07	6,550	5,700	40	0.23	17,970	11,450

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[1] ft/min x 0.3048 = m/min

Appendix B-4.3. AMMONIA REMOVAL, MLSS, RETENTION TIME AND F/M RATIO DATA SUMMARY - AS-II

Date	Influent NH_3-N, mg/l	Effluent NH_3-N, mg/l	Percent NH_3-N, removal	MLSS[a] mg/l	Hydraulic retention time, hours	Substrate loading day^{-1}
11/30/79	123	454	-	4,930	25.5	0.023
12/3	236	400	-	7,050	28.7	0.028
12/5	282	350	-	4,950	25.8	0.053
12/7	321	356	-	4,830	26.6	0.060
12/10	249	308	-	4,975	29.0	0.041
12/12	224	277	-	5,375	29.5	0.034
12/14	178	280	-	5,350	29.8	0.027
12/17	200	263	-	5,450	28.5	0.031
12/19	185	218	-	4,970	27.7	0.032
12/21	121	213	-	5,200	28.1	0.020
12/24	101	140	-	6,050	27.2	0.015
12/26	118	174	-	6,100	24.8	0.019
12/28	116	174	-	6,100	25.3	0.018
12/31	180	162	10.0	6,100	24.6	0.029
1/2/80	192	162	15.6	5,800	29.0	0.027
1/4	211	179	15.2	6,120	26.8	0.031
1/7	103	78	24.3	6,400	26.1	0.015
1/9	68	39	42.6	6,950	25.3	0.009
1/11	70	17	75.7	7,700	25.6	0.008
1/14	150	17	88.7	7,925	27.1	0.017
1/16	206	17	91.7	8,130	28.1	0.022
1/18	228	8	96.5	7,000	27.5	0.028
1/21	171	2	98.8	6,375	29.9	0.021
1/23	160	1	99.4	6,300	28.9	0.021
1/25	156	3	98.1	6,800	26.9	0.020
1/28	234	6	97.4	6,950	27.5	0.029
1/30	136	45	66.9	6,650	26.0	0.019
2/1	211	22	89.6	6,150	27.0	0.030
2/2	144	17	88.2	5,750	25.0	0.024

a. MLSS = Observed MLSS - 950 mg/l PAC

continued

Appendix B-4.3 (continued)

Date	Influent NH_3-N, mg/l	Effluent NH_3-N, mg/l	Percent NH_3-N, removal	MLSS[a] mg/l	Hydraulic retention time, hours	Substrate loading day^{-1}
2/3/80	158	22	86.1	6,050	26.5	0.024
2/4	163	22	86.5	6,350	26.4	0.023
2/5	160	11	93.1	6,350	26.3	0.023
2/6	168	1	99.4	5,950	26.1	0.026
2/7	144	2	98.6	5,600	26.1	0.024

a. MLSS = Observed MLSS - 950 mg/l PAC

Appendix B-4.4. PILOT PLANT DATA FROM THREE SELECTED OPERATING PERIODS (STAGE II)

Date	Influent NH_3-N, mg/l	Effluent NH_3-N, mg/l	MLSS[a] mg/l	Detention time, hours	Flow, ℓ/min	F/M day^{-1}	SVI	Settling rate ft/min	Effluent SS, mg/l
1/9/80	68	39	6,950	25.3	3.92	0.009	56	0.11	34
1/10	112	-	7,300	25.2	3.93	0.015	54	0.08	-
1/11	70	17	7,700	25.6	3.87	0.008	54	0.11	25
1/12	128	-	7,550	25.9	3.82	0.016	55	0.11	-
1/13	151	-	7,830	27.3	3.63	0.017	53	0.10	-
1/14	150	17	7,925	27.1	3.65	0.017	55	0.10	47
Ave.	113	24	7,543	26.1	3.80	0.014	55	0.10	35
1/16	206	17	8,130	28.1	3.53	0.022	44	0.12	56
1/17	248	-	7,750	26.1	3.80	0.029	43	0.15	-
1/18	228	8	7,000	27.5	3.61	0.028	40	0.13	66
1/19	248	-	7,250	28.3	3.51	0.029	40	0.15	-
1/20	199	-	7,300	28.8	3.44	0.023	40	0.17	-
1/21	171	2	6,375	29.9	3.31	0.021	45	0.17	66
1/22	173	-	6,600	30.8	3.21	0.020	45	0.17	-
1/23	160	1	6,300	28.9	3.43	0.021	47	0.25	51
Ave.	204	7	7,088	28.6	3.49	0.024	43	0.16	60
1/30	136	45	6,650	26.0	3.81	0.019	45	0.18	253
1/31	221	-	6,500	26.4	3.76	0.031	40	0.23	-
2/1	211	22	6,150	27.0	3.67	0.030	44	0.23	306
2/2	144	17	5,750	25.0	3.96	0.024	42	0.23	337
2/3	158	22	6,050	26.5	3.74	0.024	38	0.23	230
2/4	163	22	6,350	26.4	3.75	0.023	34	0.20	275
2/5	160	11	6,350	26.3	3.76	0.023	35	0.23	340
2/6	168	1	5,950	26.1	3.79	0.026	-	-	240
2/7	144	2	5,600	26.1	3.79	0.024	40	0.23	150
Ave.	167	18	6,150	26.2	3.78	0.025	40	0.22	266

ft/min x 0.3048 = m/min

a. MLSS = Observed MLSS - 950 mg/l PAC

Appendix B-4.5. SELECTED TEST PERIODS OF AS-2 OPERATION - FOLLANSBEE

Date	SS mg/l	VSS mg/l	NH_3-N as N mg/l	TKN as N mg/l	NO_2/NO_3-N mg/l	Phenol mg/l	CN_T mg/l	CN_A mg/l	O & G mg/l	TOC mg/l	Total P as P mg/l	$S^=$ mg/l	BOD mg/l	SCN mg/l	Color APHA Units
AS-2 Diluted Feed - Period 1															
1/09/80	11	6	68		<1		2.7	0.3		28	1.0		5	56	
10										39					
11	10	8	70	78		<0.1	1.8	1.1	2	35	1.2	0.1	5	76	150/200
12										57					
13										77					
14	34	16	150	169		<0.1	6.1	3.1	3	68	3.5		20	140	
AS-2 Effluent - Period 1															
1/09/80	34	18	39				3.1	<0.05		30	0.9		15	<1	
10										22					
11	25	13	17	22		<0.1	2.1	0.2	22	21	1.3	A	5	<1	100/150
12										24					
13										26					
14	47	16	17	20		<0.1	6.1	0.8	20	26	1.9		4	<1	
AS-2 Diluted Feed - Period 2															
1/16/80	47	23	206	221			2.0	<0.05		51	38		35	44	
17										40					
18	34	20	228	267	<1	0.1	3.6	<0.05	3	56	3.7	0.7	13	4	150/200
19										59					
20										52					
21	26	14	171	174	<1	0.7	5.6	<0.05	2	33	3.2		4	3	
22										58					
23	29	19	160	168	<1		8.8	<0.05		70	3.4		24	2	
24										65					
25	41	24	156	170		0.2	9.5	<0.05	4	64	4.1	0.9	16	1	250/300
26										59					
27										65					
28	57	30	234	266		0.1	13.3	<0.05	7	75	3.1		97	2	

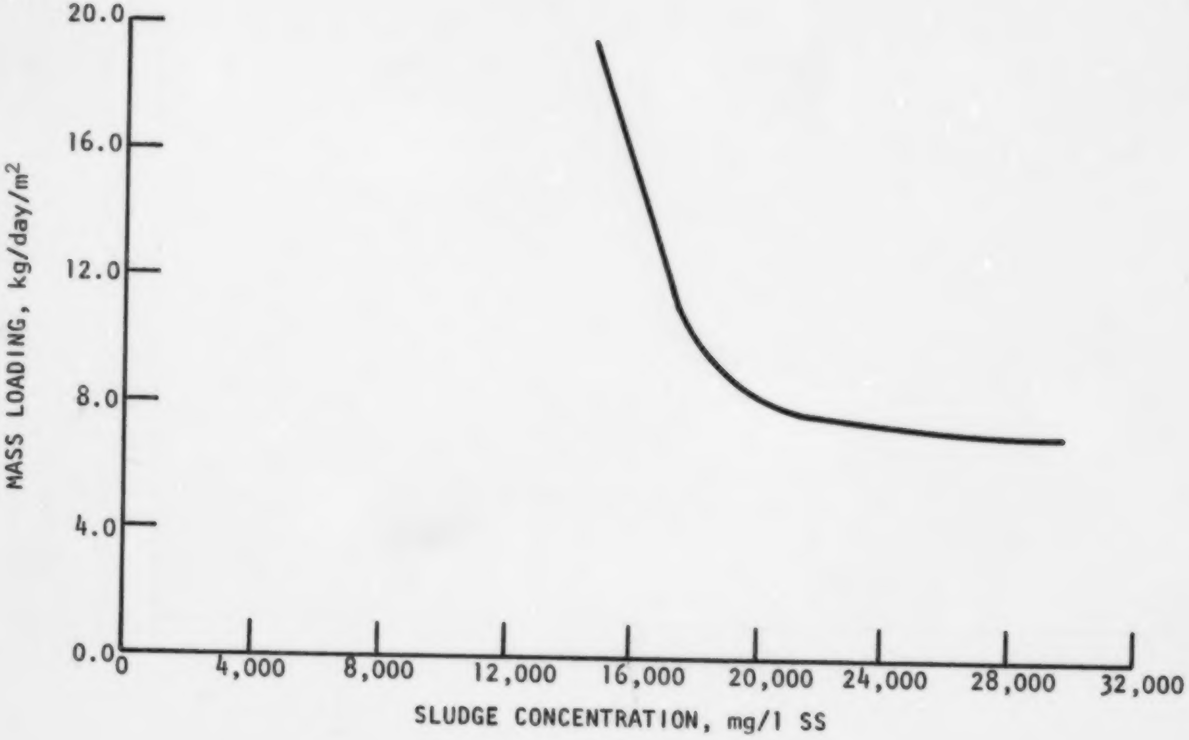
A = Interference

continued

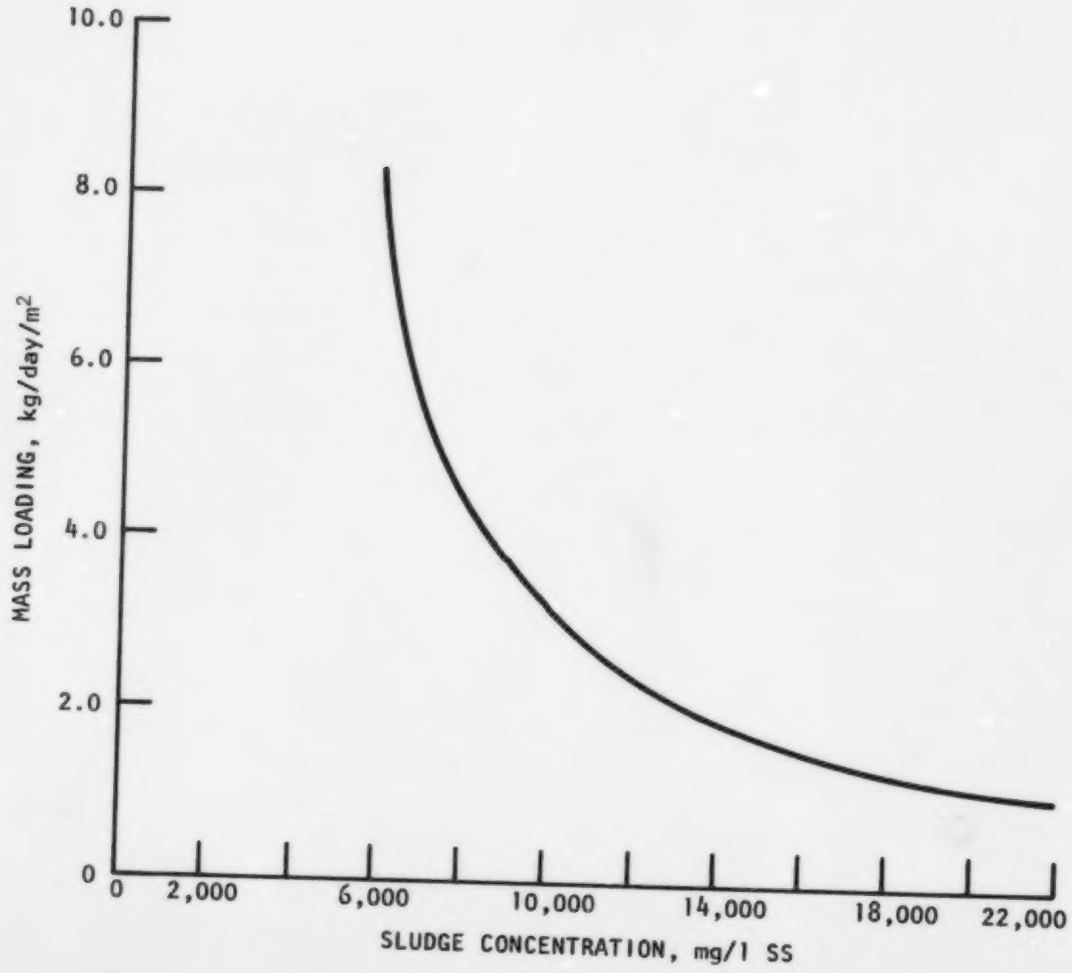
Appendix B-4.5 (continued)

Date	SS mg/l	VSS mg/l	NH_3-N as N mg/l	TKN as N mg/l	NO_2/NO_3-N mg/l	Phenol mg/l	CN_T mg/l	CN_A mg/l	O & G mg/l	TOC mg/l	Total P as P mg/l	$S^=$ mg/l	BOD mg/l	SCN mg/l	Color APHA Units
AS-2 Effluent - Period 2															
1/16/80	56	30	17	20			7.0	2.6		36	2.7		5	<1	
17										38					
18	66	41	8	11	205	<0.1	5.1	<0.05	11	51	3.7	A	7	<1	150/200
19										59					
20										62					
21	61	40	2	10	180	0.6	5.4	<0.05	8	67	3.1		6	<1	
22										72					
23	51	32	1	9	160		4.8	<0.05		104	3.6		6	<1	
24										70					
25	59	29	3	9		0.1	7.1	<0.05	6	57	3.8	A	7	<1	400/600
26										40					
27										58					
28	93	40	6	8		<0.1	8.9	<0.05	3	60	3.3		6	<1	
AS-2 Diluted Feed - Period 3															
2/01/80	70	31	211	237	<1	0.5	5.4	1.9	2	123	2.3	1.8	15	180	400/500
02	44	22	144	161	<1	<0.1	3.4	0.3	<1	31	1.6	0.8	15	105	200/300
03	58	28	158	175		0.2	1.5	<0.05	3	66	2.1	1.7	20	91	200/300
04	88	33	163	183	<1	0.1	5.3	0.9	3	57	2.4	0.7	25	31	300/400
05	95	51	160	179	<1	0.4	5.4	<0.05	7	84	2.6	0.5	55	26	300/400
06	53	44	168	170	<1	0.6	3.8	0.26	4	109	3.4	0.5	84	105	200/300
07	30	28	144	164	<1	0.3	3.2	<0.05	5	114	3.6	1.2	67	140	500/600
AS-2 Effluent - Period 3															
2/01/80	306	173	22	50	215	<0.1	8.9	5.1	5	114	2.5	A	50	<1	500/600
02	337	167	17	28	195	<0.1	9.0	<0.05	6	120	2.3	A	39	<1	400/600
03	230	142	22	42		<0.1	9.7	<0.05	12	154	2.4	A	42	<1	400/500
04	275	150	22	39	170	0.1	8.1	<0.05	12	104	2.0	A	41	<1	300/350
05	340	220	11	31	170	0.4	8.6	<0.05	21	158	2.6	A	87	<1	300/400
06	240	230	1	17	185	0.7	8.0	0.4	12	104	4.2	A	56	<1	400/500
07	150	95	2	17	190	0.9	7.0	<0.05	14	114	4.0	A	24	<1	500/600

A = Interference



Appendix B-4.6. Flux concentration curve for RS-2 on 2/2/80.



Appendix B-4.7. Flux concentration curve for RS-1 on 2/7/80.

Appendix B-5.1. DUAL MEDIA FILTRATION FIELD OPERATING DATA - BIOLOGICAL TEST SITE

Date	Run No.	Cumm. time hrs:min	Cumm. flow gal.	Filtration rate gpm/ft^2	P_{inf}, psi	P_1 psi	P_2 psi	P_3 psi	P_{eff} psi
2/2/80	1	0	0	10.0	6.0	6.0	6.0	6.0	6.0
		2:10	68.7		6.5	6.5	6.5	6.5	6.5
		3:00	95.0		6.5	6.5	6.5	6.0	5.5
		3:50	107.2		6.0	6.0	6.0	4.5	4.0
		5:45	135.5		8.0	7.5	7.0	5.0	4.0
		6:35	147.2		9.5	9.0	8.0	6.0	4.5
		6:35	147.2		Backwashed filter at 28.85 gpm/ft^2 for 10 minutes (until backwash ran clear)				
2/2/80	2	0	0	5.0	10.5	10.0	10.0	10.0	10.0
		1:10	17.8		10.0	9.5	9.5	9.0	8.5
		1:55	29.0		9.0	8.5	8.5	8.0	7.5
		3:05	46.0		9.5	9.0	8.5	8.0	7.5
		4:30	68.1		11.0	10.0	9.5	8.0	7.5
		4:30	68.1		Backwashed filter at 23.8 gpm/ft^2 for 10 minutes (until backwash ran clear)				
2/3/80	3	0	0	3.0	10.5	10.5	10.5	10.5	10.0
		3:15	26.8		10.5	10.5	10.0	9.0	8.5
		5:00	40.0		10.0	10.0	9.5	8.5	7.5
		7:00	56.3		10.0	9.5	9.0	8.0	7.5
		8:00	65.1		10.0	9.5	9.0	8.0	7.5
		8:00	65.1		Backwashed filter at 20.0 gpm/ft^2 for 10 minutes (until backwash ran clear)				
2/7/80	4	0	0	5.0	9.0	9.0	9.0	9.0	9.0
		1:30	23.5		9.5	9.5	9.5	9.0	8.0
		2:30	39.3		9.0	9.0	9.0	8.5	8.0
		4:00	64.1		11.0	10.5	9.0	7.5	5.0
		5:30	90.3		15.0	14.0	12.0	8.0	4.0
		7:30	122.8		10.0	10.0	10.0	6.0	4.0
		7:30	122.8		Backwashed filter at 18.0 gpm/ft^2 for 10 minutes (until backwash ran clear)				

(continued)

Appendix B-5.1 (continued)

Date	Run No.	Cumm. time hrs:min	SS, mg/l			VSS, mg/l			O&G, mg/l			Cu, mg/l			Cd, mg/l		
			Inf	Eff	BW	Inf	Eff	BW	Inf	Eff	BW	Inf	Eff	BW	Inf	Eff	BW
2/2/80	1	0															
		2:10															
		3:00															
		3:50															
		5:45															
		6:35	475	395		295	230		6	14							
2/2/80	2	0															
		1:10															
		1:55															
		3:05															
		4:30	205	218		205	218		12	13							
		4:30	0 min BW SS = 1770 mg/l VSS = 1700 mg/l			2 min BW SS = 1110 mg/l VSS = 950 mg/l			3 min BW SS = 91 mg/l VSS = 69 mg/l			5 min BW SS = 14 mg/l VSS = 12 mg/l			8 min BW SS = 9 mg/l VSS = 7 mg/l		
2/3/80	3	0															
		3:15															
		5:00															
		7:00															
		8:00	223	170		198	110		12	12							
		8:00															
2/7/80	4	0	187	94	5950	145	64	4750	11	10	22	0.06	0.05	0.86	0.03	0.03	0.13
		1:30															
		2:30															
		4:00															
		5:30															
		7:30															
		7:30															

(continued)

Appendix B-5.1 (continued)

Date	Run No.	Cumm time hrs: min	Pb, mg/l			Zn, mg/l			As, mg/l			Se, mg/l			Sb, mg/l		
			Inf	Eff	BW	Inf	Eff	BW	Inf	Eff	BW	Inf	Eff	BW	Inf	Eff	BW
2/2/80	1	0															
		2:10															
		3:00															
		3:50															
		5:45															
		6:35															
		6:35															
2/2/80	2	0															
		1:10															
		1:55															
		3:05															
		4:30															
		4:30															
2/3/80	3	0															
		3:15															
		5:00															
		7:00															
		8:00															
		8:00															
2/7/80	4	0	<0.1	<0.1	0.20	2.4	0.9	30.0	<0.005	<0.005	<0.005	0.56	0.42	2.33	<0.05	<0.05	<0.05
		1:30															
		2:30															
		4:00															
		5:30															
		7:30															
		7:30															

(continued)

Appendix B-5.1 (continued)

Date	Run No.	Cumm. time hrs:min	Cumm. flow gal.	Filtration rate, gpm/ft^2	P_{inf}, psi	P_1 psi	P_2 psi	P_3 psi	P_{eff} psi
2/8/80	5	0	0	8.0	11.0	10.5	10.0	10.0	10.0
		1:45	41.6		11.0	10.5	10.0	10.0	8.5
		5:00	117.8		12.5	11.5	10.0	7.0	6.0
		6:50	161.4		12.0	11.0	9.5	6.5	5.5
		9:00	212.6		12.0	11.0	9.5	6.5	5.5
		9:00	212.6		Backwashed filter at 18 gpm/ft^2 for 10 minutes (until backwash ran clear)				
2/9/80	6	0	0	10.0	10.0	10.0	10.0	10.0	10.0
		1:15	40.6		11.5	11.5	11.5	11.5	11.0
		3:30	111.6		12.0	11.5	11.5	11.5	10.0
		6:00	186.9		12.0	11.5	11.5	9.0	8.0
		7:45	239.9		12.0	11.5	11.0	7.5	6.5
		7:45	239.9		Backwashed filter at 15 gpm/ft^2 for 15 minutes (until backwash ran clear)				
2/10/80	7	0	0	8.0	12.0	12.0	12.0	12.0	12.0
		1:15	29.1		11.0	10.5	10.5	10.5	10.5
		2:15	52.6		11.0	10.5	10.5	10.5	10.5
		4:10	98.5		11.5	10.0	9.5	9.0	8.0
		6:40	157.7		11.5	10.0	9.0	8.0	7.0
					Backwashed filter at 15 gpm/ft^2 for 15 minutes (until backwash ran clear)				
2/11/80	8	0	0	3.0	10.5	10.3	10.5	10.5	10.5
		1:20	12.7		12.0	12.0	12.0	12.0	12.0
		2:45	26.2		10.5	10.5	10.5	10.5	10.5
		4:45	45.2		11.0	11.0	11.0	11.0	11.0
		6:45	64.2		11.0	10.0	10.0	10.0	10.0
		7:15	69.0		11.0	10.0	10.0	10.0	10.0

(continued)

NOTE: gpm/ft^2 × 40.74 = l/min/m^2
psi × 704 = kg/m^2

Appendix B-5.1 (continued)

Date	Run No.	Cumm. time, hrs:min	SS, mg/l			VSS, mg/l			O&G, mg/l			Cu, mg/l			Cd, mg/l		
			Inf	Eff	BW	Inf	Eff	BW	Inf	Eff	BW	Inf	Eff	BW	Inf	Eff	BW
2/8/80	5	0															
		1.45															
		5:00															
		6:50															
		9:00	137	49		113	32		8	8							
		9:00															
2/9/80	6	0															
		1:15										No polymer added during this period. See clarifier polymer data.					
		3:30															
		6:00															
		7:45	160	99		107	72		10	10							
		7:45															
2/10/80	7	0															
		1:15															
		2:15															
		4:10															
		6:40	90	61		60	48		2	4							
2/11/80	8	0															
		1:20															
		2:45															
		4:45															
		6:45															
		7:15	48	32		47	25		5	4							

(continued)

TABLE B-5.2 DUAL MEDIA FILTRATION OPERATING DATA SUMMARY - BIOLOGICAL TEST SITE

Filtration Rate		Run	Headloss		Run Time	Suspended Solids, mg/l		Percent Solids	Solids Loading	
ℓ/min/m²	gpm/ft²	No.	Meters	Feet	Hrs	Inf.	Eff.	Removal	kg SS/m²/m H.L.	lb SS/ft²/ft H.L.
		3	1.8	5.8	8.0	223	170	24%	1.760	0.110
122.2	3.0	8	0.7	2.3	7.3	48	32	33%	1.223	0.076
		Average	1.3	4.1	7.7	136	101	29%	1.492	0.093
		2	2.5	8.1	4.5	205	218	-	-	-
203.7	5.0	4	4.2	13.9	7.5	187	94	50%	2.013	0.126
		Average	3.4	11.0	6.0	196	156	-	-	-
		5	4.6	15.0	9.0	137	49	64%	3.389	0.211
325.9	8.0	7	3.2	10.4	6.7	90	61	32%	1.199	0.075
		Average	3.9	12.7	7.8	114	55	48%	2.294	0.143
		1	3.5	11.6	6.6	475	395	17%	3.651	0.228
407.3	10.0	6	3.9	12.7	7.8	160	99	38%	3.006	0.187
		Average	3.7	12.2	7.2	318	247	28%	3.328	0.208

APPENDIX C. CHILTON FACTORS USED FOR COST ESTIMATE OF THE VARIOUS TREATMENT PROCESSES

Item number	Item description	Coke plant wastewater pretreatment	Stage I activated sludge (carbonaceous removal)	Stage II activated sludge (nitrification)	Dual media filtration
(1)	Delivered equipment cost	--	--	--	--
(2)	Equipment installation	.4(1)	.4(1)	.4(1)	.4(1)
(3)	Installed equipment cost	(1)+(2)	(1)+(2)	(1)+(2)	(1)+(2)
(4)	Process pipint	.3(3)	.3(3)	.3(3)	.3(3)
(5)	Instrumentation	.05(3)	.05(3)	.05(3)	.05(3)
(6)	Capital equipment total	(3)+(4)+(5)	(3)+(4)+(5)	(3)+(4)+(5)	(3)+(4)+(5)
(7)	Building & Site Development				
	outdoor				
	outdoor/indoor	.3(6)	.3(6)	.3(6)	--
	indoor	--	--	--	.8(6)
(8)	Auxillaries (Electrical Item Power)	.2(6)	.2(6)	.2(6)	.2(6)
(9)	Other	.1(6)	.1(6)	.1(6)	.1(6)
(10)	Facilities total	(7)+(8)+(9)	(7)+(8)+(9)	(7)+(8)+(9)	(7)+(8)+(9)
(11)	Total physical plant costs	(6)+(10)	(6)+(10)	(6)+(10)	(6)+(10)
(12)	Engineering and construction	.25(11)	.25(11)	.25(11)	.25(11)
(13)	Contingency & construction	.25(11)	.25(11)	.25(11)	.25(11)
(14)	Facilities & fees	(10)+(12)+(13)	(10)+(12)+(13)	(10)+(12)+(13)	(10)+(12)+(13)
(15)	Total plant cost	(6)+(14)	(6)+(14)	(6)+(14)	(6)+(14)

(n) Indicates item number multiplied.

APPENDIX D . TYPICAL COST ESTIMATE UTILIZING CHILTON FACTORS

Item number	Item Description	Cost[a]	Comments
(1)	Delivered equipment cost	$100,000	--
(2)	Equipment installation .4x(1)	40,000	Factor used is at low end of range because some equipment is skid mounted.
(3)	Installed equipment cost	$140,000	--
(4)	Process piping .3x(3)	42,000	Factor used is at low end of range for fluid piping because some equipment is prepiped.
(5)	Instrumentation .05x(3)	7,000	Factor represents minor instrumentation.
(6)	Capital equipment total (3+4+5)	$189,000	--
(7)	Buildings & site development .3x(6)	$ 57,000	Factor is at low midrange for outdoor/indoor installation.
(8)	Auxilliaries (Electrical Power) .2x(6)	38,000	Factor is at low midrange for a major addition.
(9)	Other .1x(6)	19,000	Factor is near low end of range.
(10)	Facilities Total (7+8+9)	$114,000	--
(11)	Total Physical Plant Costs (6+10)	$303,000	Capital equipment plus facilities.
(12)	Engineering and construction fee .25(11)	76,000	Factor is for simple to moderate complexity.
(13)	Contingency & contractor's fee .25(11)	76,000	Factor is midrange for processes subject to change.
(14)	Facilities & fees (10+12+13)	$266,000	--
(15)	Total Plant Cost (6+14)	$455,000	--

(n) Indicates item number being multiplied.

[a] Rounded to the nearest thousand.